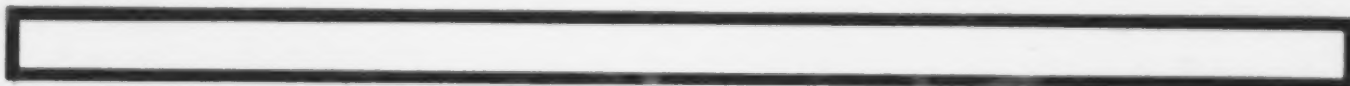


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Review of New Source Performance
Standards for Petroleum Refinery Fuel Gas

(U.S.) Environmental Protection Agency
Research Triangle Park, NC

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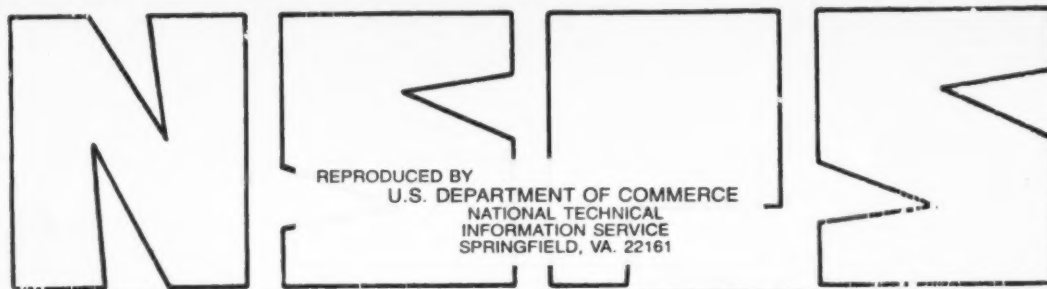
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Review of New Source Performance Standards for Petroleum Refinery Fuel Gas



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Review of New Source Performance Standards for Petroleum Refinery Fuel Gas

Emission Standards and Engineering Division

**U.S. ENVIRONMENTAL PROTECTION AGENCY
Office of Air and Radiation
Office of Air Quality Planning and Standards
Research Triangle Park, NC 27711**

October 1986

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1. SUMMARY

1.1 CONTROL TECHNOLOGY

Currently, petroleum refineries are using two types of control technologies, the alkanolamine or the LO-CAT® system, to comply with the hydrogen sulfide (H_2S) concentration limit (230 mg/dscm) of this new source performance standard (NSPS). Data for fifteen compliance tests were obtained from nine different refineries. All of the test data are from refineries that use an alkanolamine system. The compliance test results range from 2.97 mg/dscm to 119.43 mg/dscm.

No compliance test data were obtained from a LO-CAT® system; however, one refinery with such a system has installed a continuous H_2S monitor. According to information submitted by the refinery, the H_2S concentration ranges from 20 ppmv to 50 ppmv and averages 30 ppmv (NSPS is 162 ppmv).

The review did not find any demonstrated technologies for controlling emissions that achieve more control than the alkanolamine or the LO-CAT® systems.

1.2 ECONOMIC CONSIDERATIONS AFFECTING THE NSPS

Another primary issue involving review of the NSPS is the cost of controls. The cost effectiveness of controlling the H_2S concentration in refinery fuel gas was estimated for four model plant sizes at three H_2S concentrations for both types of control systems. For most of the new NSPS units, the costs of control per unit of sulfur dioxide (SO_2) removed will be less than those discussed in this section. The cost effectiveness ranges from \$201 per ton of SO_2 to a credit of \$23.9 per ton of SO_2 for the alkanolamine system with a Claus sulfur recovery unit. The cost effectiveness for the LO-CAT® system ranges from \$89 per ton to \$399 per ton of SO_2 removed.

2. CURRENT STANDARDS

This chapter presents and discusses the current regulations for air pollutant emissions from refinery fuel gas combustion devices. Federal regulations for new sources, other Federal regulations, and State regulations (for existing and new sources) are all addressed in order to give an overall picture of the regulatory structure for this emission source category. The focus, however, is on the new source performance standards (NSPS) for sulfur dioxide emissions from refinery fuel gas combustion devices.

A summary of the NSPS is first presented, followed by detailed discussions of the specific requirements, definitions, and specifications of the NSPS. This is followed by a brief description of other Federal and State regulations that may also affect existing and new sources in this category.

2.1 NEW SOURCE PERFORMANCE STANDARDS

2.1.1 Background

New source performance standards regulate emissions of air pollutants from new, modified, and reconstructed facilities in various industrial categories. The regulations establish emission limits and require emission performance testing, continuous monitoring, and periodic reporting. The authority for the NSPS regulations is granted to the U.S. Environmental Protection Agency (EPA) under Section 111 of the Clean Air Act.¹

The regulation for fuel gas combustion devices in petroleum refineries is listed in Subpart J of 40 CFR 60, (Code of Federal Regulations; Title 40 - Protection of Environment; Part 60 - Standards of Performance for New

Stationary Sources; Subpart J - Standards of Performance for Petroleum Refineries). Subpart J addresses specific requirements for this source category, but Subpart J also incorporates the general requirements for any NSPS. These general requirements are listed in Subpart A (General Provisions) of 40 CFR 60.

Other sources of air pollution emissions from petroleum refineries are also regulated under the new source performance standard regulatory program. Subpart J also regulates sulfur dioxide emissions from Claus sulfur recovery plants, and sulfur dioxide, carbon monoxide, and particulate emissions from fluid catalytic cracking unit catalyst regenerators. Subpart H regulates sulfuric acid mist and sulfur dioxide emissions from sulfuric acid plants. Volatile organic compound (VOC) emissions from storage vessels for petroleum liquids are regulated under Subparts K and Ka. Subpart GGG regulates VOC emissions due to leaks from process equipment.

2.1.2 Summary of the NSPS for Refinery Fuel Gas Combustion Devices

New source performance standards were promulgated by the EPA on March 8, 1974, limiting emissions of sulfur dioxide (SO_2) from new, modified and reconstructed fuel gas combustion devices at petroleum refineries. No significant changes have occurred since then. These standards apply to an affected facility which commences construction or modification after June 11, 1973.

The affected facility for this standard is any fuel gas combustion device in a petroleum refinery. These devices are defined as any equipment used to burn fuel gas, such as process heaters, boilers, and flares, but some combustion sources in a refinery are specifically exempted in the definition.

The regulated air pollutant is SO_2 . Sulfur dioxide emissions from fuel gas combustion devices can be controlled by reducing the hydrogen sulfide (H_2S)

content of the fuel gas prior to combustion or by flue gas desulfurization (FGD) after combustion. The standard was written with the intent to limit the H_2S content of fuel gas, although the owner/operator has the option of using FGD. The standard prohibits the burning of fuel gas containing more than 230 mg H_2S /dscm (0.10 gr/dscf) in any fuel gas combustion device; however, the standard does not apply to unusual situations, such as emergency gas releases or process upsets. Compliance is demonstrated by an initial performance test using EPA Method 11. Subsequent continuous monitoring of H_2S in the incoming fuel gas is required; however, since monitor performance specifications have not been established yet, this monitoring requirement is not in effect.

An alternative compliance option is included. Instead of controlling the H_2S in the incoming fuel gas, the SO_2 emissions may be controlled directly by treating the effluent gases resulting from the combustion of fuel gas. However, it must be shown that treating the effluent combustion gases will control SO_2 emissions as effectively as controlling the H_2S in the incoming fuel gas. Compliance for this option is demonstrated by an initial performance test using EPA Method 6, and continuous monitoring of SO_2 in the effluent gas is required.

The regulation also specifies a series of reporting and recordkeeping requirements. A refinery that has combustion devices subject to the NSPS is required to keep records, submit reports to EPA, and notify EPA of particular plans and occurrences as described in section 2.1.8.

2.1.3 Applicability of Standards²

2.1.3.1 Affected facilities.

The NSPS is applicable to any new, modified, or reconstructed combustion device which commenced construction after June 11, 1973, and which burns fuel gas in a petroleum refinery.

Petroleum is defined as,

"the crude oil removed from the earth and the oils derived from tar sands, shale, and coal."

A petroleum refinery is defined as,

"any facility engaged in producing gasoline, kerosene, distillate fuel oils, residual fuel oils, lubricants, or other products through distillation of petroleum or through redistillation, cracking or reforming of unfinished petroleum derivatives."

The EPA's definition of a petroleum refinery is thus interpreted broadly to encompass shale oil refineries, solvent refined coal plants, one-step topping plants, and small re-refining operations.

Fuel gas is defined as,

"any gas which is generated at a petroleum refinery and which is combusted. Fuel gas also includes natural gas when the natural gas is combined and combusted in any proportion with a gas generated at a refinery. Fuel gas does not include gases generated by catalytic cracking unit catalyst regenerators and fluid coking burners."

The specific exemption is included for gases generated by these particular processes because it is impractical to control the SO₂ emissions that would result from burning the H₂S in these gases. These off-gases contain relatively low levels of H₂S and contain very high levels of carbon dioxide, making it difficult to reduce the H₂S concentration further in conventional amine treating units. However, if these exempted, off-gases are combined with fuel

gas from another part of a refinery, then the combined gas stream would be subject to the NSPS limit if it is combusted in a new combustion device.

Natural gas refers to pipeline standard natural gas (meeting specifications of < 0.25 grains H_2S / 100 scf). Thus, if natural gas is purchased and burned exclusively in a combustion device, then H_2S content of the gas will necessarily be below the NSPS limit. If natural gas is mixed with refinery generated fuel gas, then the combined stream is considered "fuel gas" and its H_2S concentration must be under the NSPS limit prior to combustion in a new source.

A fuel gas combustion device is defined as,

"any equipment, such as process heaters, boilers and flares used to combust fuel gas, except facilities in which gases are combusted to produce sulfur or sulfuric acid."

There are two reasons for including the specific exemption for combustion devices used to produce sulfur or sulfuric acid. First, the combustion in this case is a step in the chemical conversion process, and the resulting post-combustion stream is considered a process stream that undergoes further processing, not an effluent stream. Second, there are separate NSPS's limiting air emissions from these processes. (Sulfuric acid plants are regulated under Subpart G; Claus sulfur recovery plants under a different part of Subpart J.)

2.1.3.2 Applicability date.

The NSPS applies only if the construction or modification commenced after June 11, 1973, (the date of the original proposal of the regulation). The term "commenced" is defined in the General Provisions to 40 CFR 60, (Section 60.2),

"Commenced means that an owner or operator has undertaken a continuous program of construction or modification or that an owner or operator has entered into a binding agreement or contractual obligation to undertake and complete, within a reasonable time, a continuous program of construction or modification."

Thus, a fuel gas combustion device that existed prior to the proposal on June 11, 1973, and has not been significantly changed or altered since then would not be regulated under the NSPS.

2.1.3.3 Modification.

While NSPS are intended primarily for newly constructed facilities, existing sources can become subject to an NSPS through either "modification" or "reconstruction." These terms are defined in detail in the General Provisions for Part 60, (40 CFR 60.14 and 40 CFR 60.15).

An existing fuel gas combustion device becomes subject to the NSPS under the modification provision if there is any physical or operational change that causes an increase in the emission rate. A number of clarifications, exemptions, and exceptions to the modification provision are listed. The following actions by themselves are not considered to be modifications:

- ° routine maintenance, repair, and replacement
- ° production increases achieved without any capital expenditure
- ° production increases resulting from an increase in the hours of operation
- ° use of an alternative fuel if the existing facility was originally designed to accommodate such an alternative use
- ° addition or replacement of equipment for emission control (as long as the replacement does not increase emissions)
- ° relocation or change of ownership of an existing facility.

Also, the addition or modification of one facility at a source will not cause other unaltered facilities at that source to become subject to the NSPS. Specifically, for petroleum refineries, if one fuel gas combustion device is added or modified, then only that particular device must meet the NSPS; the other existing combustion devices throughout the refinery are not affected.

2.1.3.4 Reconstruction

An existing facility becomes subject to the NSPS upon reconstruction regardless of any change in the rate of emissions. Reconstruction is defined as the replacement of components of an existing facility to the extent that the cumulative fixed capital cost of the new components exceeds 50 percent of the cost that would be required to construct a comparable entirely new facility.

2.1.4 Controlled Pollutant

The NSPS limits emissions to the atmosphere of SO_2 from the exhaust of refinery combustion devices which burn fuel gas. Although the regulated air pollutant is SO_2 , the SO_2 emissions to the atmosphere are not regulated directly. Instead SO_2 emissions are controlled indirectly by regulating the amount of H_2S in the incoming fuel gas.

2.1.5 Exceptions to the Emission Standard

The standard prohibits the burning of fuel gas containing more than 230 mg $\text{H}_2\text{S}/\text{dscm}$ (0.10 grain/dscf) in any new, modified, or reconstructed combustion device. The H_2S content of the incoming fuel gas can be reduced in three ways - by using sweeter crude with a lower sulfur content, by pretreatment of the fuel gas before combustion in an acid gas treating unit, and by blending natural gas with the fuel gas. Although this mixing dilutes the H_2S concentration without reducing overall SO_2 emissions, the blending of natural gas with

fuel gas is not considered a circumvention of the standard, because this is often normal and necessary refinery operating practice. Natural gas is used as auxiliary fuel since many refineries require more gas than they generate. Also, natural gas is used to ensure a constant flow of fuel to processes, while the amount of refinery generated fuel gas may vary with operating conditions or upsets in other process units.

The standard does not apply to unusual situations, such as emergency gas releases due to process upsets. Process upset gas is defined as,

"any gas generated by a petroleum refinery process unit as a result of start-up, shut-down, upset, or malfunction."³

Start-up, shut-down, upset, and malfunction considerations are defined in the General Provisions for 40 CFR Part 60. The combustion of process upset gas in a flare, or the combustion in a flare of process gas or fuel gas which is released to the flare as a result of relief valve leakage is specifically exempted from this regulation. However, flares which burn a continuous process gas stream are subject to the NSPS regulations.

2.1.6 Testing Requirements

The owner or operator of a fuel gas burning device subject to NSPS is required to conduct performance tests within a specified period after start-up, and thereafter from time to time as may be specified by the EPA. These performance tests are required in order to demonstrate that the standards are being met by the new device. General testing and reporting requirements are listed in the General Provisions, (Section 60.7), while testing details specific to this source category are found in Subpart J, (Section 60.106).

The initial test of performance of a facility must be conducted within 60 days after the facility first achieves its maximum intended rate of operation.

However, if the intended rate of operation is not achieved within 120 days of initial start-up, the initial test must nevertheless be conducted within 180 days of start-up. Thirty days must be allowed for prior notice to the EPA, to allow the Agency to designate an observer to witness the test.

To demonstrate compliance with the standard limiting the amount of H_2S in the fuel gas prior to combustion, EPA Reference Method 11 is used to determine the concentration of H_2S . A performance test consists of 3 runs, with each run consisting of 2 samples. Samples are taken at approximately one-hour intervals with a minimum sampling time of 10 minutes per sample. The arithmetic mean of the three runs constitutes the value used to determine whether the facility is in compliance. (Necessary modifications in the details of the test methods may be made, if approved in advance by the EPA.) A written report of the test is to be furnished to the EPA.

2.1.7 Monitoring Requirements

The regulation requires a continuous H_2S monitor on the fuel gas line at the inlet to each combustion device that is subject to the NSPS. When a refinery has several fuel gas combustion devices having a common source of fuel gas, monitoring may be done at one location instead of each combustion device having a separate monitor. This situation is common in many refineries where a centralized acid gas treatment plant treats H_2S rich gases from several refinery process units and then routes the treated exit gases to combustion devices throughout the refinery. Excess emissions are defined as any 3 hour period when the integrated (or arithmetic) average concentration of H_2S in the fuel gas exceeds the standard of 230 mg $H_2S/dscm$.

The monitoring system must continuously monitor and record the H₂S concentration. Under the General Provisions (Section 60.13), "continuous" is defined to mean that the monitoring system must complete at least one cycle of operation (sampling, analysis, and data recording) for each successive 15-minute period. The owner or operator must install, calibrate, maintain, and operate the continuous monitor according to the requirements which are detailed in the subpart and the general provisions. Also, the continuous monitoring system must satisfy the performance specifications in Appendix B of 40 CFR 60.

The EPA has not yet developed instrument performance specifications for H₂S continuous monitoring systems. Therefore, refinery combustion devices subject to the NSPS are effectively exempt from the H₂S monitoring requirements until EPA establishes performance specifications for an H₂S monitor.

2.1.8 Recordkeeping and Reporting Requirements

A refinery that has combustion devices subject to the NSPS, is required to keep records, submit certain reports to EPA, and notify EPA of certain plans and occurrences.

One-time "notification" reports concerning the start of construction or reconstruction, anticipated and actual startup dates, and physical or operational changes to existing facilities are required so that the EPA will be able to identify affected sources. In addition, other records and reports are necessary to enable the EPA to identify sources that may not be in compliance with the standard. These include initial performance test results, quarterly reports of excess emissions, and retaining records of plant upsets and excess emissions for 2 years. Details of these requirements are listed in the General Provisions (40 CFR 60.7).

2.1.9 Option for an SO₂ Emission Limit

Sulfur dioxide emissions from fuel gas combustion devices can be controlled by reducing the H₂S in the incoming fuel gas or by reducing the SO₂ in the exhaust gases. The standard was written with the intent to limit the H₂S content of fuel gas, and all of the refinery sources subject to the NSPS to date are currently reducing SO₂ emissions by pretreatment of the fuel gases to remove the H₂S.

However, the regulation includes a second, alternative provision which allows the burning of fuel gas with a higher H₂S content provided the effluent gases are treated to reduce the SO₂ emissions. This flue gas desulfurization would probably be accomplished with an add-on air pollution control device. In case this emission control option is selected, the regulation sets forth a parallel set of emission standards and testing, monitoring, reporting, and recordkeeping requirements. Because no refinery has yet elected the FGD options, the alternative provisions in the regulation will not be discussed in detail in this report.

The NSPS does not set a specific SO₂ emission limit if this alternative approach is followed. Instead, the source must calculate and determine an equivalent SO₂ emission level that would control SO₂ emissions as effectively as complying with the standard for H₂S concentration in the fuel gas. Because the inlet fuel gas streams, operating conditions, and parameters may vary for each combustion device, an equivalent SO₂ emission limit would probably need to be calculated for each affected facility on a case-by-case basis. The information must be submitted to the EPA for approval.

Compliance for this option is demonstrated by a performance test using EPA Methods 1, 2, and 6 for determining the sampling site, volumetric flowrate, and SO₂ concentration. Continuous monitoring of the SO₂ concentration in the effluent is also required in the regulation. Since EPA has listed performance specifications for continuous SO₂ instruments (in Performance Specification 2 of Appendix B to 40 CFR 60), these monitoring requirements are in effect and must be carried out. Excess emissions are defined as a 3-hour period when the average SO₂ concentration exceeds a predetermined level that was previously calculated and approved for that particular combustion source. The testing, monitoring, recordkeeping, and reporting requirements are similar to those discussed for the H₂S concentration standard, and details are listed in Subpart J and the General Provisions.

2.2 LEGISLATIVE HISTORY

Standards of performance for air emissions from petroleum refineries were first proposed on June 11, 1973 (38 FR 15406) and promulgated on March 8, 1974 (39 FR 9308). Since then, there have been 3 proposed and 8 final rulemakings which affected the standard for sulfur dioxide emissions from fuel gas combustion sources in refineries. These rulemakings consisted of minor clerical corrections, changes to the monitoring requirements, and changes to the definition of fuel gas. A listing and brief summary of these is given in Table 2.1.

2.2.1 Changes to Definitions

The definitions of "fuel gas" and "fuel gas combustion device" were changed to clarify the original intent of the regulation and to match the conventional nomenclature used in the industry. These formal changes to

TABLE 2.1

LIST OF RULEMAKINGS AFFECTING NSPS FOR
REFINERY FUEL GAS COMBUSTION DEVICES

DATE	FEDERAL REGISTER CITATION	TYPE OF ACTION	SUMMARY OF ACTION
06-11-73 03-08-74	38 FR 15406 39 FR 9308	Proposal Final	° Original regulation limiting SO ₂ emissions and including testing, monitoring, and reporting requirements.
09-11-74 10-06-75	39 FR 32852 40 FR 46250	Proposal Final	° Added universal monitoring and reporting requirements to General Provisions, (40 CFR 60.7 & 60.13). ° Eliminated monitoring requirements for H ₂ S in fuel gas; revised monitoring requirements for SO ₂ in exhaust gas.
07-25-77	42 FR 37936	Final	° Clarification, adding applicability date to the definition of the affected facility.
08-17-77	42 FR 41424	Final	° Clerical change to revise statutory authority citations for clarity.
03-03-78	43 FR 8800	Final	° Clerical change to amend statutory authority citations per Clean Air Act Amendments of 1977.
10-04-76 03-15-78	41 FR 43866 43 FR 10866	Proposal Final	° Re-added monitoring requirements for H ₂ S in fuel gas.
03-12-79	44 FR 13480	Final	° Change definition of "fuel gas" and "fuel gas combustion device" to clarify when an incinerator-waste heat boiler is affected by the NSPS.
03-03-80 12-01-80	45 FR 13991 45 FR 79452	Proposal Final	° Change definition of "fuel gas" to clarify which gaseous fuels are covered by NSPS, particularly when "natural gas" is a "fuel gas".

the regulation were initiated in part as a response to several questions which had arisen concerning applicability determinations. Since the changes were merely clarifying the original intent of the regulation, these changes did not have a significant impact on emissions.

2.2.2 Monitoring

The original standard included requirements for the monitoring of the H₂S at the inlet or SO₂ at the outlet of a fuel gas combustion device. At the time, however, no guidance or instructions were provided on how to conduct the monitoring, and so sources were temporarily exempted from monitoring. (Incidentally, this approach was taken for all of the early NSPS regulations covering several industries because no monitoring guidelines had been set by EPA at the time for any pollutant.)

Major changes and additions to monitoring requirements were issued a short time later (proposed 9/11/74 and promulgated 10/6/75). These changes added overall monitoring and reporting requirements to the General Provisions for all NSPS's and set forth detailed performance specifications for SO₂ monitoring instruments. Concurrently, there were coordinating revisions to the SO₂ monitoring requirements in Subpart J for the exhaust gases from refinery combustion devices. At the same time, all monitoring requirements for H₂S levels in the inlet gases were eliminated because specifications for those instruments had not yet been set. Then, this approach was reversed in a later rulemaking (proposed on 10/4/76 and promulgated on 3/15/78), when monitoring requirements for H₂S in the fuel gas were reinstated; however,

since detailed performance specifications for H₂S monitors have not been promulgated, the H₂S monitoring requirements still do not have to be carried out. The overall result of these actions and current status of requirements for H₂S monitoring is that universal, general monitoring requirements and guidelines are in place but specific, detailed requirements for H₂S monitors have not yet been determined. Thus, no H₂S monitoring is required at this time.

2.3 STATE REGULATIONS

2.3.1 State Regulations for Existing Sources

A review of State regulations has shown a wide variation in the types of regulations, degree of stringency, and methods of enforcing the limitations. Many States have several forms of regulations, each applying to a different fuel or type of source, and do not necessarily conform to or parallel the NSPS format. For example, with regard to source category, the NSPS specifies fuel gas combustion devices in petroleum refineries, whereas a State regulation may specify a different source category: petroleum refinery combustion sources; a combustion source in any industrial facility; combustion sources that burn gaseous fuels.

State regulations may limit emissions of SO₂ by limiting the H₂S in the fuel gas (as does the NSPS), the total SO₂ emissions from the combustion device, or the total SO₂ emissions from the petroleum refinery. In general, SO₂ emissions are limited by a regulation restricting the quantity of SO₂ emitted per unit quantity of heat input or by limiting the sulfur content of the fuels. In some States, the regulation specifies the maximum allowable ground level SO₂ concentration resulting from the emissions.

Sulfur dioxide regulations fall into one of the following regulatory formats:

1. ppm SO₂, by volume, in the effluent
2. Pounds of SO₂ per million Btu's of heat input
3. Requirements on the sulfur content of the fuel, such as ppm H₂S in the fuel gas or weight percent sulfur in fuel oil
4. Ambient air quality levels similar or the same as the National Ambient Air Quality Standard (NAAQS) for SO₂

2.3.2 State Regulations for New Sources

For new sources, most States have been delegated enforcement authority for the NSPS for the petroleum refining industry, and most have adopted the NSPS as written. No State has adopted an emission limit more stringent than the NSPS limit of 0.10 gr H₂S/dscf; although some States or local air pollution control agencies have adopted more stringent monitoring requirements. The Puget Sound agency (in Washington State) and three counties in southern California have required continuous emission monitoring with a continuous automated instrument even though this requirement is officially not in effect under the NSPS until instrument performance specifications are promulgated.

2.4 OTHER FEDERAL REGULATIONS

There has been additional regulatory activity since the promulgation of the current NSPS which affects the emissions from refinery fuel gas combustion sources. In addition to State regulations (for existing and new sources) and the NSPS, some petroleum refineries may be required to achieve more stringent emission levels from fuel gas combustion devices under regulations for the Prevention of Significant Deterioration (PSD)⁴ or under State Implementation Plans (SIP) which are subject to nonattainment review by EPA.⁵

2.4.1 PSD Regulations

If a new facility is built in an area which is attaining the NAAQS for

SO₂, then it falls under the PSD regulations and must use the best available control technology (BACT). In general, BACT determination is applied on a case-by-case basis and must be at least as stringent as the NSPS level. For this source category, the BACT emission level has been defined as equal to the NSPS level.⁶ Some State and local enforcing agencies have also used the PSD regulations as a means of requiring some form of emission monitoring and corresponding emission monitoring reports.

2.4.2 Nonattainment Area Regulations

If a new source is located in a nonattainment area for NAAQS for SO₂, then emission control technology capable of the lowest achievable emission reduction (LAER) is required. In general, LAER is at least as stringent as the NSPS, but for this source category, LAER has also been defined as equal to the NSPS.⁷

2.4.3 Other NSPS Regulations

The NSPS under review in this study limits the SO₂ emissions from all new, modified, or reconstructed fuel gas combustion devices, which include process heaters, boilers, and flares used to combust fuel gas. The air pollution emissions from these combustion sources may also be regulated by other NSPS's (either current or under development). The additional requirements and restrictions imposed by these other NSPS's do not affect or conflict with the emission reductions required by the fuel gas NSPS, but they may affect the planning and design of new combustion sources by refinery owners.

A new or modified boiler which has a heat input greater than 250 million BTU per hour may also be subject to the provisions of Subpart D or Subpart Da of 40 CFR 60; the applicable subpart is based on the date of construction of the source and whether the source is classified as an industrial boiler

or a utility boiler. These subparts limit opacity, particulate emissions, and NO_x emissions, as well as SO₂ emissions. Similarly, a medium-sized, new or modified boiler (100 to 250 million BTU per hour) may be subject to the provisions of Subpart Db, which is currently under development.

2.5 References

1. Clean Air Act As Amended, August 1977. 42 U.S.C. Title I--Air Pollution Prevention and Control. Part A--Air Quality and Emission Limitations; Section 111--Standards of Performance for New Stationary Sources. Washington, D.C.
2. U.S. Environmental Protection Agency. Code of Federal Regulations. Title 40, Part 60. Sections 60.101. Office of the Federal Register. Washington, D.C. July 1, 1985.
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5. Reference 1. Part D--Plan Requirements for Nonattainment Areas.
6. U.S. Environmental Protection Agency. BACT/LAER Clearinghouse - A Compilation of Control Technology Determinations. U.S. EPA, Research Triangle Park, N.C. EPA 450/3-85-016 (a-d). June 1985.
7. Same as reference #6.

3. INDUSTRY DESCRIPTION

This chapter describes a typical petroleum refinery, its various production processes and the range of products produced by a refinery. Also, the current number of operating refineries and their geographical location are discussed as well as the industry growth rate. Finally, emissions from combustion of refinery fuel gas and the rational for choosing refinery fuel gas for new source performance standard control are discussed.

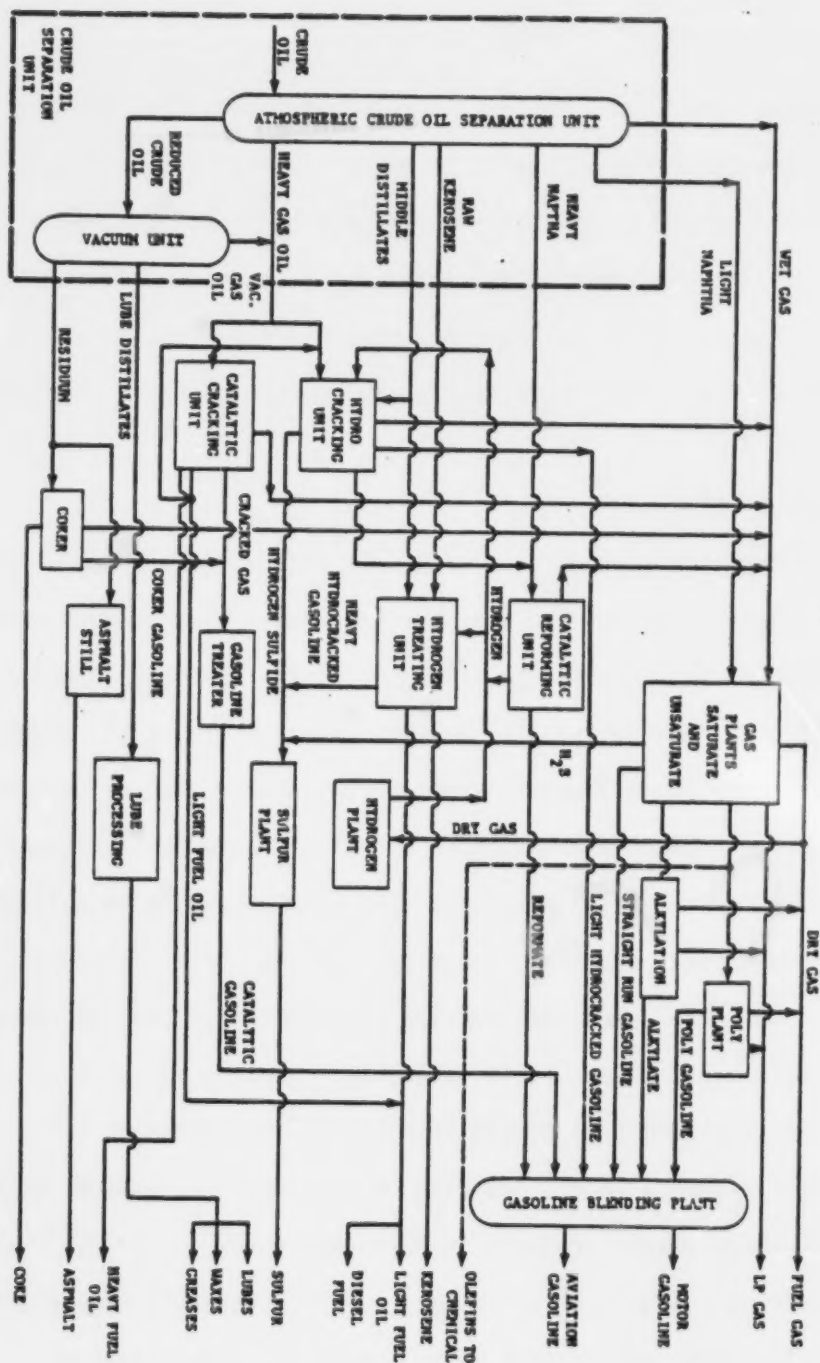
3.1 BACKGROUND INFORMATION

A petroleum refinery transforms crude oil into a variety of useful products. The petroleum refining industry produces more than 2500 products that can be categorized into the following classes: fuel gas, gasoline, kerosene, fuel oil, lubricating oil, grease, wax, asphalt, coke, chemicals, and solvents. There is no "typical" refinery, since the number of products and the product mix varies widely within a refinery as well as between refineries. The manufacturing processes also vary depending on refinery age, type of technology, capacity, location, and type of crude processed.

Petroleum refinery operations involve physical separation of components of the crude oil (e.g., crude distillation) and chemical conversion processes which transform some of the less useful components of the oil into more useful products (e.g., cracking of high molecular weight oils into lower molecular weight products such as gasoline).

The processing sequence of a refinery is illustrated in Figure 3-1. The crude oil is heated and charged to an atmospheric distillation

FIGURE 3-1



tower where it is separated into several light, intermediate, and heavy fractions. The bottoms from the tower are sent to a vacuum distillation unit for further separation. The bottoms from the vacuum still are thermally cracked in a coker to produce a wet gas, coker gasoline, and coke. A portion of the bottoms from the vacuum still may be processed into asphalt. Gas oils from the atmospheric and vacuum distillation units are used as feedstocks for the catalytic cracking and hydrocracking units. These units convert the gas oils to gasoline and distillate fuel. The gasoline from these units is fed to a catalytic reformer to improve the octane number and then blended with other refinery streams to make gasolines for marketing.

The wet gas streams from the distillation, coker, and cracking units are combined and fractionated into fuel gas, liquified petroleum gas, and unsaturated and saturated branched chain and straight chain, light hydrocarbons containing from three to five carbon atoms. The fuel gas is used as fuel in the refinery furnaces. The straight chain saturated hydrocarbons are blended into gasoline. The unsaturated hydrocarbons and the branched chain hydrocarbons, primarily isobutane, are processed in an alkylation unit. In the alkylation unit the unsaturated hydrocarbons react with isobutane to form isoparaffins which are blended into gasoline to increase the octane.

The middle distillates from the crude unit, the coker unit, and the cracking unit are blended into diesel and jet fuels and furnace oil. Heavy vacuum gas oils and reduced crude oil from some crudes can be processed into lubricating oils, waxes, and grease.

3.2 INDUSTRY CHARACTERIZATION

As of January 1, 1985, there were 191 operable refineries in the United States with total crude oil distillation capacity of 15.1 million barrels per calendar day^{a/} and 15.9 million barrels per stream day.^{b/} Table 3.1 lists the number of operating refineries along with the total crude capacity located in each State. These refineries are distributed among 35 states, with 79 refineries (41 percent) being concentrated in the three major refining States of Texas, California, and Louisiana. These three States account for an even higher percentage (57 percent) of the total U.S. crude oil refining capacity. Texas alone accounts for approximately 27 percent of the total crude oil refining capacity, while Louisiana and California account for about 15 percent each.

In addition, there are four operating refineries in Puerto Rico, Virgin Islands, Guam, and the Hawaiian Foreign Trade Zone with a combined crude capacity of 771,000 barrels per calendar day.² Because these are territories and not formally part of the United States, these refineries are usually not included in industry studies and figures for the U.S.; however, these refineries are regulated under federal EPA new source performance standards.

^{a/} Barrels per calendar day (b/cd): the average volume a refinery unit processes each day including downtime used for turnarounds. This is actual total volume for the year divided by 365.

^{b/} Barrels per stream day (b/sd): the amount a unit can process running at full capacity under optimal crude and product slate conditions for short periods.

TABLE 3.1
OPERATING REFINERIES IN THE U.S. (AS OF JANUARY 1, 1985)³

State	No. Plants	Crude Capacity	
		b/cd	b/sd
Alabama	1	80,000	81,300
Alaska	4	138,930	142,910
Arizona	1	5,000	5,263
Arkansas	4	69,170	70,950
California	30	2,265,098	2,381,417
Colorado	3	94,700	98,500
Delaware	1	140,000	150,000
Georgia	2	28,800	32,000
Hawaii	2	109,500	118,426
Illinois	8	946,000	1,003,550
Indiana	5	431,300	445,500
Kansas	7	338,000	352,383
Kentucky	2	218,900	226,000
Louisiana	16	2,188,793	2,280,958
Maryland	1	14,200	14,947
Michigan	4	119,400	126,094
Minnesota	2	204,143	211,220
Mississippi	5	362,400	383,104
Montana	6	147,500	154,147
Nevada	1	4,500	4,700
New Jersey	5	503,000	533,210
New Mexico	3	63,050	66,000
North Dakota	2	62,800	65,400
Ohio	5	515,700	540,000
Oklahoma	5	374,000	390,394
Oregon	1	15,000	15,789
Pennsylvania	8	658,700	691,300
Tennessee	1	57,000	60,000
Texas	33	4,145,900	4,385,273
Utah	6	154,950	161,868
Virginia	1	51,000	53,000
Washington	7	410,550	427,543
West Virginia	2	16,500	17,000
Wisconsin	1	39,000	40,000
Wyoming	6	162,778	168,052
Total	191	15,136,262	15,898,198

Since January 1, 1981, a net total of 101 refineries have been shut down with a total capacity of 2.5 million barrels per calendar day. (Table 3-2 and Table 3-3). During 1984, a total of 26 refineries were shut down and 2 refineries were started up resulting in a net decrease of 24 plants with an associated loss in crude distillation capacity of 0.5 million barrels per calendar day. The majority of these closings occurred at refineries with crude distillation capacity of 30,000 barrels per calendar day or less. These closings accounted for a net reduction of 19 facilities. Refineries with crude distillation capacity greater than 30,000 barrels per calendar day showed a net reduction of 5 facilities during 1984.

Refinery utilization (actual production vs. production capacity) peaked at 78.2 percent in August 1984; the average rate for the year rose to 76.2 percent, from the previous year's average of 71.7 percent.⁴

Total downstream charge capacity on January 1, 1985, stood at 28.3 million barrels per stream day, a net decrease of 0.4 million barrels per stream day. Downstream charging capacity includes the following processes: vacuum distillation, thermal operation, catalytic cracking (fresh and recycled), catalytic reforming, catalytic hydrocracking, and catalytic hydrotreating. New construction at existing refineries and the start-up of previously closed refineries was more than offset by the nearly 0.7 million barrels per stream day capacity closed during 1984. The most significant declines were in vacuum distillation and catalytic reforming, dropping 2 percent and 4 percent, respectively. However, during the year, downstream charge capacity increased 3 percent for catalytic cracking (recycled) and nearly 11 percent for catalytic hydrocracking.⁵

TABLE 3-2

Number of Operable Refineries in the U.S. 6,7,8,9,10

<u>Year</u>	<u>Total</u>
1981	303
1982	273
1983	225
1984	220
1985	191

TABLE 3-3

Crude Oil Distillation 11,12,13,14,15,16

(Thousands of Barrels per Calendar Day)

	<u>1981</u>	<u>1982</u>	<u>1983</u>	<u>1984</u>	<u>1985</u>
U.S. Capacity	18,465	17,669	16,157	15,862	15,136
Puerto Rico	228	255	244	121	121
Virgin Islands and Guam	714	744	559	558	588
Hawaiian Foreign Trade Zone	68	68	60	60	62

Refiners project that total downstream charge capacity may drop to 28.1 million barrels per stream day by 1986. However, the downstream processing mix is not projected to change appreciably from January 1, 1985, levels. All major downstream refinery processes, except catalytic hydrotreating, are expected to fall below current year levels.¹⁷

Refinery receipts of crude oil averaged 12.2 million barrels per day during 1984, up 0.5 million barrels per day from 1983. This increase represented a reversal of the downward trend that started in 1979. Receipts of domestic crude oil averaged 8.8 million barrels per day and foreign averaged 3.3 million barrels per day. Most of the increase in refinery receipts of crude oil was from domestic sources, rising from 8.6 million barrels per day during 1983 to 8.8 million barrels per day during 1984. Foreign crude receipts also rose during 1984, reaching 3.3 million barrels per day from 3.2 million barrels per day during 1983.¹⁸

3.3 EMISSIONS FROM COMBUSTION OF REFINERY FUEL GAS

An integrated refinery uses energy equivalent to about 10 percent of the total energy content in the crude oil it processes.¹⁹ The fuel requirements of any one refinery depends on the nature of the feed, the final product yield and the level of the individual product quality. All of the refinery's energy needs could be derived from its own crude oil feed, usually refinery fuel gas and residual oil, but most refineries are designed to use available supplemental fuels such as natural gas.

Figure 3-1 illustrates a modern refinery, the various processes and their respective products. As indicated in Figure 3-1, several refinery processes produce refinery fuel gas as a by-product. After removing the

H₂S, the refinery fuel gas is burned in various combustion devices (boilers and heaters) located throughout the refinery. Based on five plant surveys and trip reports, there do not appear to be any particular processes or combustion devices in which refineries utilize refinery fuel gas as a fuel. The number of combustion devices and the various processes in which refinery fuel gas is burned varies greatly from one refinery to another.

Since all crude oil contains some amount of sulfur, the refinery fuel gas produced by the various processes will also contain sulfur. The sulfur content of crude oil ranges from less than 0.1 percent to greater than 5 percent sulfur by weight. Sulfur in the refinery fuel gas will be in the form of hydrogen sulfide (H₂S), carbonyl sulfide, mercaptan sulfur plus C₁ and C₂ sulfides. As the sulfur concentration of the crude oil increases, so does the concentration of H₂S and other sulfur compounds in the refinery fuel gas. Combustion of refinery fuel gas containing H₂S produces sulfur dioxide (SO₂). Thus, when untreated refinery fuel gas derived from crude oil with a high sulfur content is burned in the various process combustion sources, substantial quantities of SO₂ will be emitted to the atmosphere.

As discussed in Chapter 2, the new source performance standard prohibits the burning of refinery fuel gas containing more than 230 mg H₂S/dscm (0.10 gr/dscf) in any combustion device that burns refinery fuel gas. The combustion of process upset gas in a flare, and process gas or refinery fuel gas released to a flare from relief valve leakage is exempt from this standard.

The alternative to the 230 mg H₂S/dscm refinery fuel gas standard is that an owner or operator may elect to treat the gases resulting from the

combustion of refinery fuel gas so as to limit the release of SO_2 to the atmosphere. The EPA Administrator must be satisfied that treatment of the combustion gases controls SO_2 emissions as effectively as compliance with the H_2S standard.

The standard is equivalent to a SO_2 content of approximately 20 gr/100 scf of refinery fuel gas burned. Burning such fuel will result in a concentration of 15 to 20 parts per million SO_2 in the combustion products.²⁰

3.4 SELECTION OF REFINERY COMBUSTION DEVICES FOR NSPS CONTROL

Combustion devices that burn refinery fuel gas were originally selected for NSPS development because of their potential to emit sulfur dioxide (SO_2) in significant quantities. At the time of the NSPS development, (in the early 1970's) the nationwide emissions of SO_2 were estimated to be 28 million tons per year. It was estimated that in 1970 approximately 0.8 million tons of sulfur dioxide were emitted from petroleum refineries.²¹ The background study for the original NSPS predicted an overall emissions reduction for controlled sources of 95 percent.

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2. National Petroleum Refiners Association. U.S. Refining Capacity. Washington, D.C. July 1, 1985.
3. Cantrell, A. Annual Refining Survey. Oil and Gas Journal. Page 122. March 18, 1985.
4. Same as reference #2. Page iii.
5. Same as reference #2. Page iii.
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7. Cantrell, A. Annual Refining Survey. Oil and Gas Journal. Page 130. March 22, 1982.
8. Cantrell, A. Annual Refining Survey. Oil and Gas Journal. Page 130. March 21, 1983.
9. Cantrell, A. Annual Refining Survey. Oil and Gas Journal. Page 112. March 26, 1984.
10. Same as reference #3. Page 123.
11. Same as reference #6.
12. Same as reference #7.
13. Same as reference #8.
14. Same as reference #9.
15. Same as reference #3. Page 123.
16. Same as reference #2.
17. Same as reference #2. Page V.
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19. 1985 HPI Market Data. Hydrocarbon Processing. page 26.
20. Background Information for Proposed New Source Performance Standards: Asphalt Concrete Plants, Petroleum Refineries, Storage Vessels, Secondary Lead Smelters and Refineries, Brass or Bronze Ingot Production Plants, Iron and Steel Plants, and Sewage Treatment Plants. U.S. Environmental Protection Agency. Research Triangle Park, N.C. Publication Number APTD-1352a. June 1973. Page 27.
21. National Air Pollutant Emission Estimates, 1940-1983. U.S. EPA, RTP, N.C. EPA 450/4-84-028. December 1984.

4. STATUS OF CONTROL TECHNOLOGY

This chapter discusses the control techniques being used to meet the new source performance standards (NSPS). In order to comply with the hydrogen sulfide (H_2S) emission requirements (230 mg/dscm, 0.10 gr/dscf) of this NSPS, the owner/operator of the affected facility has the option of either reducing the H_2S concentration prior to burning the refinery fuel gas in the affected facility or treating the sulfur dioxide (SO_2) emissions from the affected facility. At this time, all known affected facilities control SO_2 emissions by reducing the H_2S concentration in the fuel gas prior to being burned.

A review of the literature, the Environmental Protection Agency (EPA) compliance data system, discussions with refinery personnel, trade organizations, local, State, and EPA regional agencies reveals that two processes are being used to comply with the NSPS. These two systems are the alkanolamine process system and the LO-CAT® system. These processes are discussed below.

4.1. ALKANOLAMINE PROCESS SYSTEM

The first commercially available alkanolamine was triethanolamine (TEA) which was used in early natural gas treating plants. As other members of the alkanolamine family were introduced into the market, they were evaluated as possible acid gas absorbents.¹ Alkanolamines are categorized as being primary, secondary, or tertiary, depending upon the degree of substitution of the central nitrogen atom by organic groups. Structural formulas for the various alkanolamines are presented in Figure 4-1. Two commercially utilized primary amines are monoethanolamine (MEA) and diglycolamine (DGA),

each shows single substitution of organic groups at the central nitrogen atom. Two commercially available secondary amines are diethanolamine (DEA) and diisopropanolamine (DIPA). The two secondary amines show double substitution of organic groups at the central nitrogen atom. A triple substitution of organic groups at the central nitrogen atom is possible, hence the tertiary amines: triethanolamine (TEA) and methyldiethanolamine (MDEA).

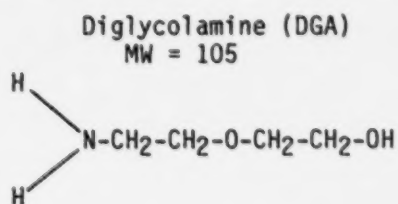
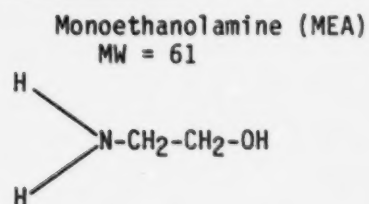
Alkanolamines are weak organic bases with each one of the amines illustrated in Figure 4-1 having at least one hydroxyl group (OH) and one amino group. In general, it can be considered that the hydroxyl group serves to reduce the vapor pressure and increase the water solubility, while the amino group provides the necessary alkalinity in water solutions to cause the absorption of acidic gases.² As crude oil is processed the following acid gases are formed and can be found in untreated refinery fuel gas: hydrogen sulfide (H_2S), carbon dioxide (CO_2), and carbonyl sulfide (COS). These compounds are considered to be acid gases because when dissolved in an aqueous medium, they dissociate to form weak acids. The acid gas and amine base will combine chemically to form an acid-base complex called a salt, thus removing the acid gas from the process stream.³

The principal reactions of the alkanolamines with H_2S are listed in Table 4-1. The reactions shown in Table 4-1 proceed to the right at low temperatures and to the left at higher temperatures. This is the reason that H_2S can be absorbed by alkanolamine solutions at ambient temperatures. At elevated temperatures (as exist in the stripper column), the reactions are reversed with the sulfide and carbamate salts being decomposed and the acid gases released in the stripper column.⁴

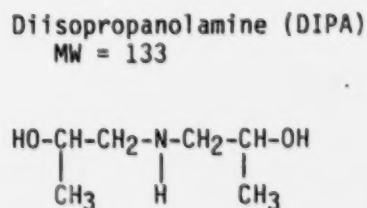
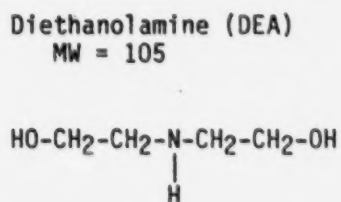
FIGURE 4-1

MOLECULAR STRUCTURES OF COMMON GAS TREATING AMINES⁵

Primary Amines: single substitution of organic group at the nitrogen atom



Secondary Amines: double substitution of organic group at the nitrogen atom



Tertiary Amines: triple substitution of organic group at the nitrogen atom

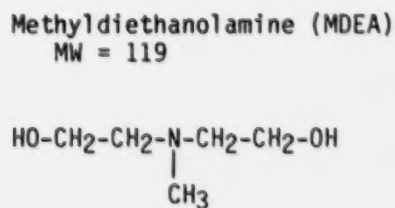
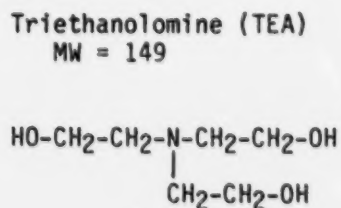


TABLE 4-1

PRINCIPAL REACTIONS OF ALKANOLAMINES WITH HYDROGEN SULFIDE⁶Primary AminesSecondary AminesTertiary Amines

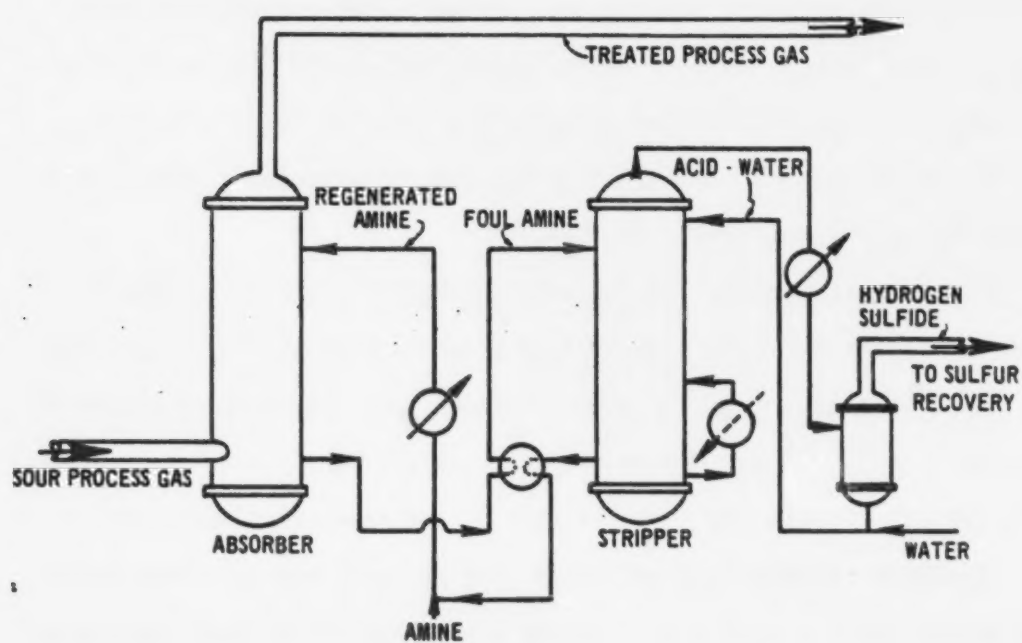
R = C₂H₄OH

Various literature sources indicate that as alkanolamines were developed and field tested, the primary and secondary amines displaced TEA. Triethanolamine was displaced largely because of its low H_2S capacity (resulting from higher equivalent weight), its low reactivity (as a tertiary amine), and its relatively poor stability.⁷ Information obtained via plant visits and contacts with petroleum refineries indicate that primary and secondary amines are indeed the principle types of amines used in alkanolamine process systems. However, one refinery responded that it used MDEA, a tertiary amine. Also, two of the five refineries that were visited are considering switching from DEA to MDEA because of its selectivity for removing only H_2S and the associated reduction in energy needed to regenerate the MDEA.^{8,9}

The general process flow for an alkanolamine processing plant is shown in Figure 4-2. The process flow scheme varies little, regardless of the aqueous amine solution used to remove H_2S . The primary pieces of equipment of concern are the absorber column and stripper column, together with the associated piping, heat exchange, and separation equipment.

The sour refinery fuel gas containing H_2S will nearly always enter the alkanolamine process plant through a separator (not shown) to remove any free petroleum liquids and/or entrained solids. The sour gas then enters the bottom of the absorber column and flows upward through the absorber in intimate counter-current contact with the aqueous amine solution. The H_2S is removed and sweetened gas leaves the top of the absorber and flows to another separator (not shown) to remove any amine solution and/or entrained solids that are carried over with the exiting refinery gas.

FIGURE 4-2
ALKANOLAMINE TREATING UNIT ¹⁰



Lean amine solution from the bottom of the stripper column is pumped through an amine-amine heat exchanger and then through a water or air-cooled exchanger before being introduced to the top tray of the absorber column. The amine moves downward through the absorber counter-current to the sour gas, and absorbs H_2S from the gas stream. Rich amine solution flows from the bottom of the absorber through the amine-amine heat exchanger and then to the top of the stripper column.

The amine-amine heat exchanger serves as a heat conservation device and lowers total heat requirements for the process. The rich solution flows downward through the stripper in counter-current contact with vapor generated in the reboiler. The reboiler vapor (primarily steam) strips the acid gases from the rich solution. The acid gases and steam leave the top of the stripper and pass through a condenser, where the major portion of the steam is condensed and cooled. The acid gases are separated in a separator and sent to the sulfur recovery unit. The condensed steam is returned to the top of the stripper column as reflux or a slip stream is bled off to control the ammonia concentration in the top of the stripper column to help maintain the amine system water balance.

Rich amine solution leaves the bottom of the absorber at an elevated temperature due to the heat of reaction released when acid gases react with the amine. The amine cooler serves to lower the lean amine temperature to the 100°F range. Higher temperature on the lean amine solution will result in excessive amine losses through vaporization and also lower acid gas carrying capacity in the solution because of temperature effects.

4.2 THE LO-CAT® HYDROGEN SULFIDE OXIDATION PROCESS

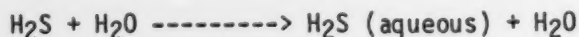
The LO-CAT® process, a process based on a liquid-phase-oxidation technique originated by Humphreys and Glasgow (London), was developed to provide an isothermal process for carrying out the modified Claus reaction.¹¹



The LO-CAT® process has two basic designs: aerobic, which absorbs H₂S from air laden streams, and anaerobic, used when the gas stream has little or no oxygen present. The anaerobic application is used in treating refinery fuel gas. Figure 4-3 is a flow diagram of an anaerobic LO-CAT® H₂S oxidation process used to treat refinery fuel gas.

The LO-CAT® process removes sulfur by using a proprietary catalyst that consists of an aqueous solution of chelated iron, buffered with soda ash (Na₂CO₃), potassium hydroxide (KOH) or any other common alkali to a pH of about 8. The LO-CAT® catalyst solution is circulated in a closed loop between the absorbers and the oxidizer vessel. Sour gas passes through an inlet gas scrubber to remove entrained liquids and enters the absorber at line pressure through a specially designed venturi prescrubber. The gas passes upward through a low liquid-filled vessel, an absorber, in which the sour gas contacts the LO-CAT® catalyst solution flowing downward. In the absorbers the H₂S gas is absorbed very rapidly into the catalyst solution where it is immediately oxidized to precipitate elemental sulfur. The following chemical reactions take place in the absorber vessel:¹²

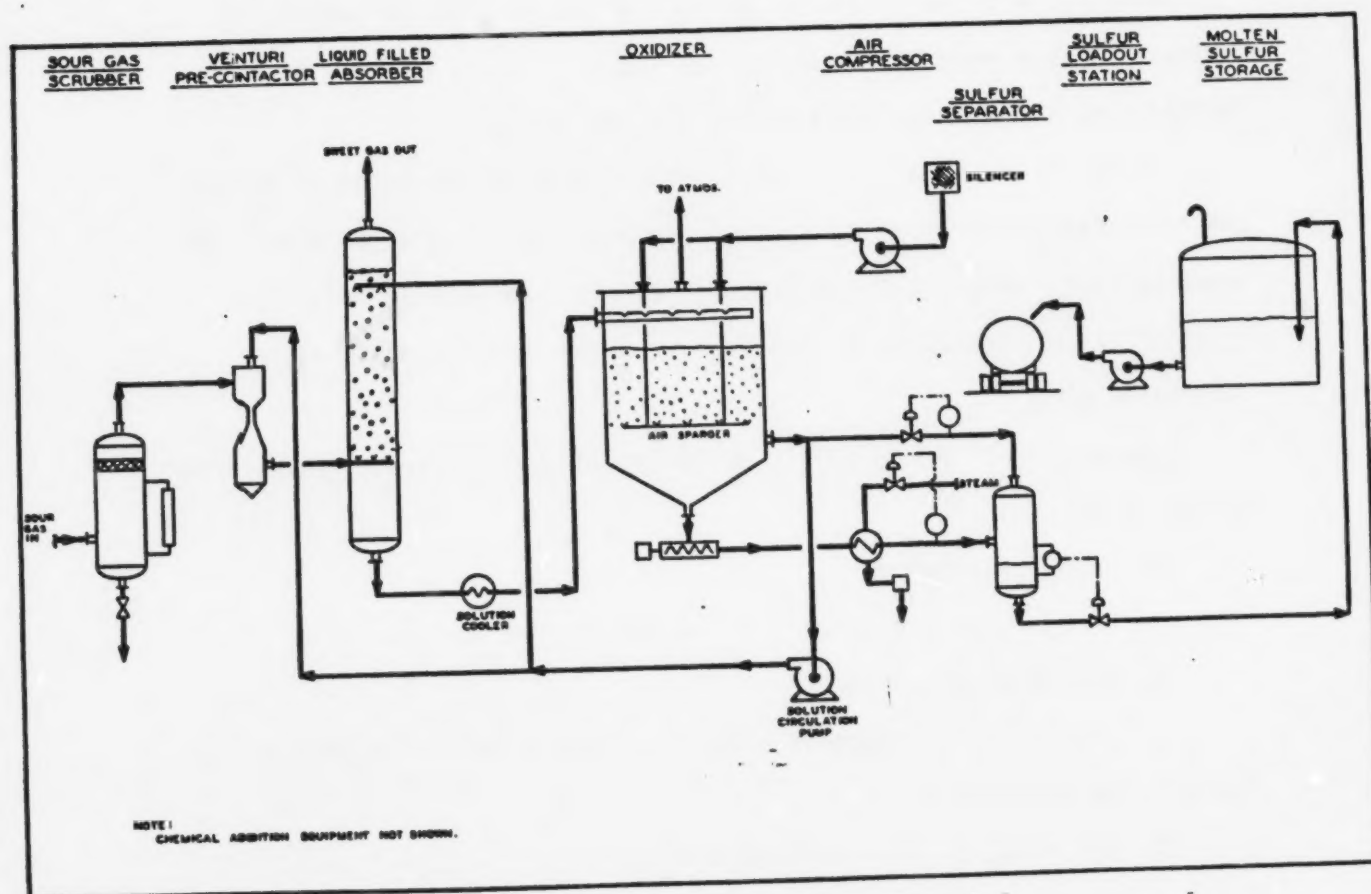
(1) H₂S Absorption



(2) First Ionization



FIGURE 4-3
FLOW DIAGRAM OF A LO-CAT SYSTEM¹³



(3) Second Ionization



(4) Oxidation of Sulfide

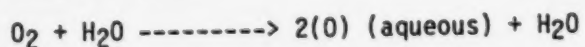


Sweet gas with less than 100 ppm H_2S leaves the absorber vessel and passes through a sweet gas scrubber (not shown in Figure 4-3) to guard against mist carryover before entering the fuel system.

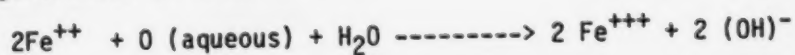
Active LO-CAT® catalyst solution enters the absorber column at the top and flows downward through the vessel counter-current to the gas flow. The spent solution leaves at the bottom of the vessel. Partially reduced solution is regenerated in the oxidizer vessel by direct contact with compressed air.

Regeneration of the LO-CAT® solution in the oxidizer vessel involves the following reactions:¹⁴

(5) Oxygen Absorption



(6) Regeneration of Iron



Overall, the reaction is

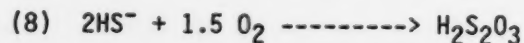


The air used to generate the catalyst solution is supplied by an air blower to an air sparger grid near the bottom of the oxidizer. The spent air, with a small fraction of the oxygen removed, leaves the top of the oxidizer vessel and discharges to the atmosphere.

The circulating catalyst solution, which contains elemental sulfur in a slurried form, flows from the absorber back to the oxidizer through a solution cooler which removes heat generated by the exothermic reactions in the absorber. Sulfur formed in the absorber vessel circulates with the solution and the particles grow to the 10-20 micrometer range. The larger particles settle out of the bulk solution, flow in the bottom of the oxidizer vessel, and are flushed out the oxidizer vessel as a slurry of 10 to 20 percent by weight of sulfur. The sulfur slurry is pumped at approximately 100 psig through a special non-plugging heat exchanger or melter, where it is heated to 270°F by 50 psig steam. The aqueous catalyst solution/molten sulfur mixture formed in the melter is fed through steam jacketed piping to a steam jacketed separator vessel. Molten sulfur is withdrawn from the separator vessel to a molten sulfur storage tank or a sulfur pit where it is stored for ultimate shipment by truck or rail car.

The clarified catalyst solution leaves the sulfur settler through a back pressure control valve set at about 75 psig. The bulk of the solution is returned to the settler vessel and then to the absorber/oxidizer circuit.

In the oxidation of H_2S to sulfur, some side reaction takes place which may be represented by the equation:¹⁵



This reaction reduces the pH of the scrubbing solution, and Na_2CO_3 , KOH , or other alkaline salt must be added to maintain the pH of the solution in the 8-8.5 pH range. This leads to a gradual buildup of sodium thiosulfate ($\text{Na}_2\text{S}_2\text{O}_3$) or similar water-soluble sulfur-containing salts in the solution. These have no deleterious effect at concentrations below 30 percent by weight. When the thiosulfate concentration exceeds 30 percent by weight

some of the spent catalyst solution is withdrawn to a holding tank for disposal. The thiosulfate has a commercial value; thus, this solution can be sold and the thiosulfate reclaimed. Another means of disposal is to treat this solution in the refinery's biological wastewater treatment system.

4.3 FACILITIES SUBJECT TO THE NSPS

In Chapter 2 the term "affected facility" was discussed and defined. As noted, the affected facility is defined as the combustion device that is capable of burning refinery fuel gas and not the H₂S control system. Information obtained from plant trips and surveys indicates that refinery fuel gas is burned in combustion devices that are used in a wide variety of production processes within a refinery. There does not appear to be a tendency for refineries to use refinery fuel gas as an energy source in any particular production process. Thus the utilization of refinery fuel gas will vary from one refinery to another.

According to information supplied by the EPA's Stationary Source Compliance Division (SSCD), fifty-four (54) petroleum refineries have become subject to the H₂S portion of this NSPS since 1975. Unfortunately, an accurate number of affected facilities within each refinery could not be determined. The information supplied by SSCD is considered to underestimate the number of refineries and affected facilities subject to this NSPS. This low estimate is attributed to the fact that some State agencies that have authority to enforce the Clean Air Act do not report all the refineries and affected facilities to SSCD.

Trade journals were also reviewed for information. These sources report new refinery construction projects plus modification and reconstruction

projects that are occurring at petroleum refineries. Unfortunately, the manner in which the modification and reconstruction projects are reported does not indicate whether these projects will be subject to the H₂S portion of this NSPS nor how many combustion sources are involved. A review of various trade journals from 1980 to 1985 indicates there have been construction projects at 42 petroleum refineries that involve either an amine treater unit, a LO-CAT® unit, or unspecified refinery fuel gas recovery.

4.4 COMPLIANCE TEST RESULTS

The Environmental Protection Agency regional offices, State agencies, and petroleum refineries were contacted to obtain compliance test information for new, modified, or reconstructed facilities.

All of the test data that were submitted were from refineries that use an alkanolamine process system to remove the hydrogen sulfide (H₂S) from the refinery fuel gas. No compliance test data were obtained from a refinery that uses the LO-CAT® system to treat sour refinery fuel gas.

4.4.1. Alkanolamine Process System Compliance Data^{16,17,18,19,20,21,22,23}

Data for fifteen (15) compliance tests were obtained from nine (9) different refineries. The results of these tests are listed in Table 4-2. The new source performance standard (NSPS) for H₂S is 230 mg/dscm (0.10gr/dscf). Compliance test results for the H₂S concentration in refinery fuel gas ranges from 2.97 mg/dscm to 119.43 mg/dscm. The data indicate that refineries are able to meet the NSPS. Also, none of the refineries that were contacted indicated that they were experiencing any problems complying with the NSPS.

TABLE 4-2
COMPLIANCE TEST RESULTS FOR ALKANOLAMINE PROCESS SYSTEMS

<u>Refinery</u>	<u>Date of Test</u>	<u>Type of Alkanolamine Used</u>	<u>Avg. H₂S Concentration (mg/dscm)</u>
A	2/81	MEA/primary	101
	4/82	" "	32
	10/82	" "	12
	3/84	" "	57
	7/84	" "	77
B	7/76	DEA/secondary	119.43
C	8/84	MEA/primary	2.97
	8/84	" "	8.09
D	11/80	DEA/secondary	12.5
E	11/81	DEA/secondary	89.2
	1/84	" "	63.3
F	--	MEA/primary	105.6
G	10/84	MEA/primary	18.1
H	8/81	DEA/secondary	27.7
I	6/85	DGA/primary	3.1

NSPS = 230 mg/dscm

4.4.2 LO-CAT® Process System²⁴

Currently, the LO-CAT® treatment system is used in only two (2) refineries within the United States to remove H₂S from refinery fuel gas. Only one of these refineries was operating as of January 1985. Both of these LO-CAT® systems are small units, less than 20 LT/D. Compliance tests have not been performed for either of the two LO-CAT® systems. Thus, there are no data for this type of system. However, one refinery has installed a continuous H₂S monitor. According to information submitted by the refinery, the H₂S concentration of the refinery gas treated by the LO-CAT® system ranges from 20 ppmv to 50 ppmv and averages 30 ppmv. The refinery reported no excess emissions have occurred since the LO-CAT® system became operational.

4.5 Emission Monitoring

The NSPS requires an H₂S continuous monitoring device to be installed in order to determine compliance with the H₂S standard of 230 mg/dscm. However, the EPA has not yet developed performance specifications for H₂S continuous monitoring systems.

In April 1979, the EPA initiated work to establish specifications and also to determine the durability, maintenance requirements, and data validity of commercially available H₂S continuous emission monitors.²⁵ Five (5) H₂S monitors were selected for evaluation. Selection criteria included operating principles, engineering judgment about suitability for use at petroleum refineries, and total cost. The performance of the five (5) H₂S monitors tested was disappointing. The absolute agreement between reference Method 11 and all monitors was poor and variable in eight out of ten relative accuracy tests. Thus, a conclusion of the test program

was that the use of H₂S monitors for compliance purposes could not be recommended at that time nor could performance specifications for H₂S monitors be written. Even though performance specifications for continuous H₂S monitoring devices have not been promulgated by the EPA, some refineries have been required by State agencies to install an H₂S continuous monitoring device.

Refineries subject to the H₂S portion of this NSPS that do not have a continuous H₂S monitor are using manually collected grab samples (Draeger Tubes) to test the refinery fuel gas for H₂S concentration. The frequency with which these samples are collected varies from one refinery to another and can range from 1 manual sample per shift to 3 manual samples per shift.

4.6 REFERENCES

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5. MODEL PLANTS AND CONTROL SYSTEMS

5.1 MODEL PLANTS AND CONTROL SYSTEMS

Model plants and model plant parameters are selected to represent the range of facilities that have become subject to the NSPS since its promulgation in 1974 or that are likely to be constructed, modified, or reconstructed in the future. The control systems applied to the model plants are selected to represent those control techniques that have been commonly installed to meet the NSPS. The model plants are defined in this chapter and a cost analysis of these model plants is presented in Chapter 6.

As discussed in Chapter 4, two types of control systems are commonly used by refineries to control H_2S in refinery fuel gas. These two systems are amine treater/Claus sulfur recovery units and LO-CAT® units.

The amine treater/Claus sulfur recovery unit combination is the most widely used process. The LO-CAT® system is a relatively new process that has been introduced since the last review of this standard and currently only two LO-CAT® units (sulfur capacity 7 and 17.35 LT/D) have been installed in the U.S. for H_2S removal from refinery fuel gas.

5.1.1 Model Control Systems and H_2S Concentrations

The selection and sizing of model H_2S control systems is based on the total sulfur loading in the untreated refinery fuel gas. Six sizes of control units were evaluated for this review: 1, 2, 5.1, 10.2, 50.8, and 101.6 Mg/D (1, 2, 5, 10, 50, and 100 LT/D). Three H_2S concentrations in the untreated refinery fuel gas (1, 5, and 10 percent H_2S by volume) were analyzed for each model plant.

Tables 5-1 and 5-2 list the model plants. The amine treater/Claus sulfur recovery unit was modeled for the following sizes: 5.1, 10.2, 50.8, and 101.6 Mg/D (5, 10, 50, and 100 LT/D). The LO-CAT® system was modeled for 1, 2, 5.1, and 10.2 Mg/D (1, 2, 5, and 10 LT/D).

5.2 EMISSION REDUCTIONS

This section presents the sulfur dioxide (SO_2) emission reductions that are achieved by each model control system. The emission reduction for each model control system is determined by comparing the emissions from a control system meeting the new source performance standard (NSPS) versus the emissions from the baseline situation of no control.

5.2.1 Amine Treater/Claus Sulfur Recovery Unit

The SO_2 emission reductions for the model amine/Claus sulfur recovery units are presented in Table 5-3. All incoming sulfur (in the form of H_2S) in the untreated fuel gas is either routed to fuel gas combustion devices, converted in the Claus sulfur recovery unit to elemental sulfur, or emitted from the Claus unit to the atmosphere. The model amine treaters were designed to achieve the NSPS limit of 230 mg of $\text{H}_2\text{S}/\text{dscm}$ in the fuel gas stream to combustion sources, while the remainder of the H_2S is sent to the Claus unit. The model Claus units were designed to achieve a 96.6 percent sulfur recovery efficiency. For calculating emission reduction, it was assumed that the amine treater/Claus system was operating at the capacity of the Claus unit for 350 operating days per year.

TABLE 5-1
MODEL PLANTS: AMINE TREATER AND CLAUS SRU

Model Plant Size Sulfur Processed (Mg/D)	(LT/D)	Fuel Gas H ₂ S Conc. (% by Volume)	Amine Efficiency (%)	Claus SRU Efficiency (%)
5.08	(5)	1	98.38	96.6
		5	99.68	
		10	99.84	
10.16	(10)	1	98.38	96.6
		5	99.68	
		10	99.84	
50.8	(50)	1	98.38	96.6
		5	99.68	
		10	99.84	
101.6	(100)	1	98.38	96.6
		5	99.68	
		10	99.84	

TABLE 5-2
MODEL PLANTS: LO-CAT®

Model Plant Size Sulfur Processed (Mg/D)	(LT/D)	RFG H ₂ S Conc. (% by Volume)	Unit Efficiency (%)
1.016	(1)	1	98.38
		5	99.68
		10	99.84
2.032	(2)	1	98.38
		5	99.68
		10	99.84
5.08	(5)	1	98.38
		5	99.68
		10	99.84
10.16	(10)	1	98.38
		5	99.68
		10	99.84

TABLE 5-3
 SO₂ EMISSION REDUCTION USING A
 AMINE TREATER/CLAUS SULFUR RECOVERY UNIT

Model System (Mg/D)	Control Size ^a (LT/D)	Uncontrolled SO ₂ Emissions ^{b,c} (Mg/yr)	SO ₂ Emission Reductions ^d (Mg/yr)
5.08	(5)	3556	3435
10.16	(10)	7112	6870
50.8	(50)	35560	34350
101.6	(100)	71120	68700

^a Model control system size (i.e., amount of sulfur processed) is based on the sulfur loading into the amine treater/Claus sulfur recovery unit.

^b 1 year = 350 operating days

^c 1 Mg of sulfur (S₂) converts to 2 Mg of SO₂ emissions

^d Efficiency of the sulfur recovery unit was assumed to be 96.6%

5.2.2 LO-CAT® Unit

The SO₂ emission reductions for the model LO-CAT® units are presented in Table 5.4. As indicated in Figure 4-3, the LO-CAT® process has no emission stream to the atmosphere. All the incoming sulfur (in the form of H₂S) is either converted to elemental sulfur or routed to the fuel gas combustion devices. The model LO-CAT® units were designed to achieve the NSPS limit of 230 mg of H₂S/dscm in the outlet fuel gas stream. For calculating emission reductions, it was assumed that the LO-CAT® unit was operating at capacity for 350 operating days per year.

TABLE 5-4
SO₂ EMISSION REDUCTION USING A LO-CAT® UNIT

Model Control System Size ^a (Mg/D)	(LT/D)	H ₂ S Conc (% by Vol.)	Uncontrolled SO ₂ Emissions ^{b,c} (Mg/yr)	Control System SO ₂ Emissions (Mg/yr)	SO ₂ Emission Reductions (Mg/yr)
1.016	(1)	1	712	11.5	701
		5	712	2.3	710
		10	712	1.1	711
2.032	(2)	1	1422	2.3	1399
		5	1422	4.6	1417
		10	1422	2.3	1420
5.08	(5)	1	3556	57.6	3498
		5	3556	11.4	3545
		10	3556	5.7	3550
10.16	(10)	1	7112	115.2	6997
		5	7112	22.8	7089
		10	7112	11.4	7101

^a Model control system size (i.e. amount of sulfur processed) is based on the sulfur loading into the LO-CAT® unit.

^b 1 year = 350 operating days

^c 1 Mg of sulfur (S₂) converts to 2 Mg of SO₂ emissions

6. COST ANALYSIS

6.1 INTRODUCTION

This section presents costs (in January 1985 dollars) of model plant control systems necessary to meet the hydrogen sulfide (and consequently the sulfur dioxide) provisions of the current NSPS for refinery fuel gas. Two control systems are analyzed: (1) for large sources of refinery fuel gas, an amine treating section coupled with a Claus sulfur recovery plant; (2) for smaller sources, a LO-CAT® process which absorbs the hydrogen sulfide and oxidizes it to elemental sulfur with the aid of an iron-based catalyst.

(A more detailed discussion of the control systems appears in Chapter 4.)

Capital and annualized costs are estimated for the following model plant sizes, all given on a nominal sulfur basis: LO-CAT® -- 1.016 Mg per day (1 long ton per day), 2.032 Mg/D (2 LT/D), 5.08 Mg/D (5 LT/D), and 10.16 Mg/D (10 LT/D); amine-Claus -- 5.08 Mg/D (5 LT/D), 10.16 Mg/D (10 LT/D), 50.8 Mg/D (50 LT/D), 101.6 Mg/D (100 LT/D), 203.2 Mg/D (200 LT/D), 508 Mg/D (500 LT/D) and a plant expansion from 10.16 Mg/D to 10.668 Mg/D (10 to 10.5 LT/D)*.

(A more detailed discussion of model plants is presented in Chapter 5.)

Cost-effectiveness is calculated for all model plants and is discussed for the two model plant sizes that are common to both control systems. Except for the plant expansion, all estimated costs apply to new control systems installed at new sources of refinery fuel gas. Note that the cost of Claus tail gas treating is not considered in the analysis.

The costs presented for the amine-Claus process are based on raw cost data provided in References 1 and 2. Costs for the LO-CAT® process are based on information from References 3, 4, and 5. Capital costs are on a

*Note: 1 Mg/D = 1.1025 short tons/D = 0.9844 LT/D

turnkey basis and thus include the purchase cost of equipment and auxiliaries, taxes, freight, and all necessary installation costs, as well as indirect costs such as engineering and supervision, construction and field expense, contractor fee, and contingency. Annualized costs include direct operating costs such as operating labor, maintenance labor, utilities, and materials, as well as indirect costs such as capital charges, overhead, property taxes, insurance, and administration. Net annualized cost is also presented, representing total annualized cost less the credit for recovered sulfur where applicable. The costs thus obtained are "order-of-magnitude"--i.e., nominally ± 50 percent in accuracy. Because most of the operating and maintenance (O&M) costs have been calculated directly, their accuracy should approach that of a study estimate (± 30 percent). Finally, because the annualized costs are comprised of O&M costs and capital cost components, their accuracy should fall between ± 30 and ± 50 percent.

6.2 AMINE TREATMENT WITH CLAUS SULFUR RECOVERY

6.2.1 Capital Costs

The capital costs for this combination process are estimated for five model plants--four new and one expansion of an existing plant. (Detailed model plant parameters are given in Chapter 5.) The capital costs for the amine treatment section were calculated from cost factors given in Reference 1 for gas sweetening operations. The process employs diethanolamine (DEA) as the absorbent.

The capital costs for the Claus section were developed by updating the costs in Reference 2 via the Chemical Engineering (CE) Plant Cost Index. The capital cost for the 5.08 Mg/D Claus plant was extrapolated logarithmically from the 10.16 Mg/D model plant cost in Reference 1 (the smallest size) using a 0.2 exponent, in accordance with cost vs. capacity formulas presented

in Appendix A of Reference 1. The 0.2 exponent was found to hold in the low size range, based on data gathered for Claus plants in Reference 1. Costs for the 50 and 100 LT/D plants were updated from costs in Reference 1 for these sizes.

Capital costs for the two sections were added together for each model plant and are presented in Table 6-1. The capital cost of the plant expansion was calculated as the incremental cost of a 10.668 Mg/D plant over a 10.16 Mg/D plant and does not include any retrofit costs. This approach was taken as an expedient to arriving at an order-of-magnitude cost for the expansion; it must be recognized that such an expansion would be a site-specific case, the cost of which could vary significantly.

The capital costs for 200 and 500 LT/D model plants can be estimated by multiplying the cost of the 100 LT/D plant by a factor of two or five, respectively. This results from the fact that plants above 100 LT/D capacity are normally constructed as trains, so that their costs vary linearly with size.

6.2.2 Annualized Costs

The annualized costs associated with owning and operating the amine-Claus plants are estimated for each model plant. The direct operating cost includes operating and supervisory labor, maintenance, supplies, utilities, chemicals, and catalysts. For the amine section, direct operating costs were calculated by means of the cost factors given in Appendix A. For the Claus section, the direct operating costs were developed by combining consumption data with unit cost data, both from Reference 2. (The latter are listed in Table 6-2.) The unit cost data were updated slightly by use of Bureau of Labor Statistics factors in most cases, such as the Producer Price Index.

TABLE 6-1
CAPITAL COST SUMMARY
AMINE TREATMENT WITH CLAUS SULFUR RECOVERY
(January 1985 Dollars)

Cost in Thousands of Dollars			
Plant, Capacity, Mg/D (LT/D) Sulfur			
	Amine ^a	Claus ^b	Total ^c
5.08 Mg/D (5 LT/D)			
1% H ₂ S	110	2290	2390
5% H ₂ S	88.8	2290	2370
10% H ₂ S	84.6	2290	2370
10.16 Mg/D (10 LT/D)			
1% H ₂ S	220	2630	2840
5% H ₂ S	178	2630	2800
10% H ₂ S	169	2630	2790
50.8 Mg/D (50 LT/D)			
1% H ₂ S	1100	4480	5570
5% H ₂ S	888	4480	5360
10% H ₂ S	846	4480	5320
101.6 Mg/D (100 LT/D)			
1% H ₂ S	2200	6470	8670
5% H ₂ S	1780	6470	8240
10% H ₂ S	1690	6470	8160
Expansion of 0.508 Mg/D (0.5 LT/D) ^d			
1% H ₂ S	10.7	25.8	36.5
5% H ₂ S	9.2	25.8	35.0
10% H ₂ S	8.5	25.8	34.3

a Cost developed from information in Reference 1.

b Cost developed from information in Reference 2.

c Totals may not add due to rounding.

d These costs represent the incremental costs of a 10.688 Mg/D plant over a 10.16 Mg/D plant.

TABLE 6-2

ASSUMED UNIT COSTS FOR
OPERATING COST ESTIMATION
CLAUS SULFUR RECOVERY PLANTS^{a,b,c}

<u>Operating Cost Item</u>	<u>Unit Cost (Credit)</u>	<u>English Equivalent</u>
Operating labor	\$15.00/h	-
Supervision labor	\$20.00/h	-
4300 Kp steam	\$17.50/Mg ^d	\$ 7.94/10 ³ lb
1760 Kp steam	(\$16.25/Mg)	(\$ 7.37/10 ³ lb).
352 Kp steam	(\$14.00/Mg)	(\$ 6.35/10 ³ lb)
106 Kp steam	(\$11.00/Mg)	(\$ 4.99/10 ³ lb)
Steam condensate	(\$ 3.00/Mg)	(\$ 1.36/10 ³ lb)
Boiler feedwater	\$ 3.60/Mg	\$ 1.63/10 ³ lb
Electricity	\$15.28/Gj	\$ 0.055/KWh
Fuel gas	\$ 4.00/Gj	\$ 4.22/MM BTU
Sulfur	(\$110/Mg) ^e	(\$111.8/LT)

- a For Claus plant operating cost estimation, unit costs were multiplied by consumption factors given in Reference 2.
- b Unit costs were not used for operating cost estimation for the amine process or LO-CAT® process; operating costs for these processes were given in References 1 and 3, respectively, as functions of gas flowrate or plant capacity.
- c The total costs for catalysts given in Reference 2 were multiplied by 1.1 to update them to January 1985 dollars.
- d Cost or credit depending upon model plant.
- e Based on the cost given for recovered (i.e., Claus plant) sulfur, Houston terminal, in Chemical Marketing Reporter, January 21, 1985.

Most indirect costs were factored from capital costs or direct operating costs using appropriate factors from References 6 and 7. Capital recovery was calculated from the capital cost with a 10 percent rate of return and a 15-year equipment life (Reference 2). There is no waste disposal cost, because spent catalysts are regenerated. The annualized cost includes an estimate of the credit for recovered sulfur at prices applicable in January 1985, a period of relative stability for recovered sulfur prices. Note also that credit for different pressures of steam varies in that, in general, the larger the Claus plant, the greater the heat recovery practiced and the higher the pressure of steam that is recovered.

The annualized costs for both sections were added together for each model plant and are presented in Tables 6-3 through 6-9. Again, costs for expanding the existing 10.16 Mg/D plant to 10.668 Mg/D were calculated as incremental costs. All annualized costs are based on 350 operating days per year. See Appendix B for an example of annualized cost estimation for the individual processes that make up the control system (amine treating and Claus sulfur recovery).

6.3 LO-CAT® PROCESS

6.3.1 Capital Costs

Capital costs for LO-CAT® model plants are also estimated. Although the model plants envision incoming hydrogen sulfide concentrations of 1, 5, and 10 percent, according to Reference 5 the differences in capital cost among the three would not be significant. Therefore, the costs are developed independent of hydrogen sulfide concentration. The purchased equipment cost was taken directly from Reference 3 and includes all primary and auxiliary equipment and instrumentation. Taxes and freight costs were obtained from Reference 6.

Table 6-3

ANNUALIZED COST SUMMARY^a
 AMINE TREATMENT WITH CLAUS SULFUR RECOVERY
 5.08 Mg/D (5 LT/D) PLANT
 (January 1985 Dollars)

Cost in Thousands of Dollars

	1% H ₂ S	5% H ₂ S	10% H ₂ S
Direct Operating Cost			
Operating labor	210	210	210
Supervision	61	61	61
Maintenance, repairs	70	70	70
Supplies, laboratory	17	17	17
Utilities			
4300 Kp steam	4	4	4
Boiler feedwater	12	12	12
Electricity	57	57	57
Fuel Gas	10	10	10
Chemicals, catalyst	3	3	3
Indirect Operating Cost			
Overhead (50 percent of all labor costs)	153	153	153
Capital Recovery (10 percent rate of return, 15 years equipment life)	315	312	312
Taxes (one percent of capital cost)	24	24	24
Insurance (one percent of capital cost)	24	24	24
Administration (two percent of capital cost)	48	47	47
Total Annualized Cost Without Credit	1,007	1,003	1,002
Credits			
4300 Kp steam	0	0	0
1760 Kp steam	48	48	48
352 Kp steam	0	0	0
106 Kp steam	3	3	3
Steam Condensate	5	5	5
Sulfur	189	189	189
Total Credits	244	244	244
Net Annualized Cost	762	759	758

^a Totals may not add due to rounding

Table 6-4

ANNUALIZED COST SUMMARY^a
AMINE TREATMENT WITH CLAUS SULFUR RECOVERY
10.16 Mg/D (10 LT/D) PLANT
 (January 1985 Dollars)

Cost in Thousands of Dollars

	1% H ₂ S	5% H ₂ S	10% H ₂ S
Direct Operating Cost			
Operating labor	210	210	210
Supervision	61	61	61
Maintenance, repairs	82	82	82
Supplies, laboratory	17	17	17
Utilities			
4300 Kp steam	7	7	7
Boiler feedwater	24	24	24
Electricity	114	114	114
Fuel Gas	19	19	19
Chemicals, catalyst	7	7	7
Indirect Operating Cost			
Overhead (50 percent of all labor costs)	155	155	155
Capital Recovery (10 percent rate of return, 15 years equipment life)	374	369	367
Taxes (one percent of capital cost)	28	28	28
Insurance (one percent of capital cost)	28	28	28
Administration (two percent of capital cost)	57	56	56
Total Annualized Cost Without Credit	1,184	1,177	1,175
Credits			
4300 Kp steam	0	0	0
1760 Kp steam	95	95	95
352 Kp steam	0	0	0
106 Kp steam	6	6	6
Steam Condensate	9	9	9
Sulfur	378	378	378
Total Credits	489	489	489
Net Annualized Cost	695	688	686

^a Totals may not add due to rounding

Table 6-5

ANNUALIZED COST SUMMARY^a
AMINE TREATMENT WITH CLAUS SULFUR RECOVERY
50.8 Mg/D (50 LT/D) PLANT
 (January 1985 Dollars)

Cost in Thousands of Dollars

	1% H ₂ S	5% H ₂ S	10% H ₂ S
Direct Operating Cost			
Operating labor	210	210	210
Supervision	61	61	61
Maintenance, repairs	151	151	151
Supplies, laboratory	17	17	17
Utilities			
4300 Kp steam	58	58	58
Boiler feedwater	170	170	170
Electricity	512	512	512
Fuel Gas	97	97	97
Chemicals, catalyst	35	35	35
Indirect Operating Cost			
Overhead (50 percent of all labor costs)	169	169	169
Capital Recovery (10 percent rate of return, 15 years equipment life)	733	705	700
Taxes (one percent of capital cost)	56	54	53
Insurance (one percent of capital cost)	56	54	53
Administration (two percent of capital cost)	111	107	106
Total Annualized Cost Without Credit	2,436	2,400	2,393
Credits			
4300 Kp steam	0	0	0
1760 Kp steam	464	464	464
352 Kp steam	18	18	18
106 Kp steam	26	26	26
Steam Condensate	50	50	50
Sulfur	1,889	1,889	1,889
Total Credits	2,448	2,448	2,448
Net Annualized Cost	(11)	(47)	(55)

^a Totals may not add due to rounding

Table 6-6

ANNUALIZED COST SUMMARY^{a,b}
AMINE TREATMENT WITH CLAUS SULFUR RECOVERY
101.6 Mg/D (100 LT/D) PLANT
 (January 1985 Dollars)

Cost in Thousands of Dollars

	1% H ₂ S	5% H ₂ S	10% H ₂ S
Direct Operating Cost			
Operating labor	210	210	210
Supervision	61	61	61
Maintenance, repairs	227	227	227
Supplies, laboratory	17	17	17
Utilities			
4300 Kp steam	0	0	0
Boiler feedwater	441	441	441
Electricity	1,007	1,007	1,007
Fuel Gas	194	194	194
Chemicals, catalyst	71	71	71
Indirect Operating Cost			
Overhead (50 percent of all labor costs)	184	184	184
Capital Recovery (10 percent rate of return, 15 years equipment life)	1,140	1,084	1,073
Taxes (one percent of capital cost)	87	82	82
Insurance (one percent of capital cost)	87	82	82
Administration (two percent of capital cost)	173	165	163
Total Annualized Cost Without Credit	3,897	3,825	3,810
Credits			
4300 Kp steam	307	307	307
1760 Kp steam	101	101	101
352 Kp steam	322	322	322
106 Kp steam	52	52	52
Steam Condensate	39	39	39
Sulfur	3,779	3,779	3,779
Total Credits	4,600	4,600	4,600
Net Annualized Cost	(702)	(775)	(789)

^a Totals may not add due to rounding.

^b Note: For estimating the cost of plants with capacities in excess of 100 LT/D, the costs in this table would be used as costs of individual trains. For example, a 200 LT/D plant would consist of two 100 LT/D trains, and the costs for the plant would be twice the costs in this table.

Table 6-7

ANNUALIZED COST SUMMARY
 AMINE TREATMENT WITH CLAUS SULFUR RECOVERY
 EXPANSION OF 10.16 Mg/D (10 LT/D) PLANT TO 10.668 Mg/D (10.5 LT/D)
 (January 1985 Dollars)

1% H₂S

	10.66 Mg/D	10.16 Mg/D	Incremental
Direct Operating Cost	210,000	210,000	0
Operating labor	60,900	60,900	0
Supervision	83,000	82,070	930
Maintenance, repairs	16,800	16,800	0
Supplies, laboratory	172,300	164,100	8,200
Utilities			
Chemicals, catalyst	7,268	6,922	346
Indirect Operating Cost			
Overhead (50 percent of all labor costs)	155,300	155,300	0
Capital Recovery (10 percent rate of return, 15 years equipment life)	379,600	374,800	4,775
Taxes (one percent of capital cost)	28,870	28,500	370
Insurance (one percent of capital cost)	28,870	28,500	370
Administration (two percent of capital cost)	57,740	57,000	740
Total Annualized Cost Without Credit	1,200,600	1,184,900	15,700
Credits			
4300 Kp steam	0	0	0
1760 Kp steam	100,100	95,360	4,740
352 Kp steam	0	0	0
106 Kp steam	6,601	6,287	314
Steam Condensate	9,767	9,302	465
Sulfur	396,800	377,900	18,900
Total Credits	513,268	488,849	24,419
Net Annualized Cost	687,330	696,050	(8,720)

Table 6-8

ANNUALIZED COST SUMMARY
 AMINE TREATMENT WITH CLAUS SULFUR RECOVERY
EXPANSION OF 10.16 Mg/D (10 LT/D) PLANT TO 10.668 Mg/D (10.5 LT/D)
 (January 1985 Dollars)

5% H₂S

	10.66 Mg/D	10.16 Mg/D	Incremental
Direct Operating Cost			
Operating labor	210,000	210,000	0
Supervision	60,900	60,900	0
Maintenance, repairs	83,000	82,070	930
Supplies, laboratory	16,800	16,800	0
Utilities	172,300	164,100	8,200
Chemicals, catalyst	7,268	6,922	346
Indirect Operating Cost			
Overhead (50 percent of all labor costs)	155,300	155,300	0
Capital Recovery (10 percent rate of return, 15 years equipment life)	373,850	369,250	4,600
Taxes (one percent of capital cost)	28,430	28,080	350
Insurance (one percent of capital cost)	28,430	28,080	350
Administration (two percent of capital cost)	56,860	56,160	700
Total Annualized Cost Without Credit	1,193,140	1,177,660	15,480
Credits			
4300 Kp steam	0	0	0
1760 Kp steam	100,100	95,360	4,740
352 Kp steam	0	0	0
106 Kp steam	6,601	6,287	314
Steam Condensate	9,767	9,302	465
Sulfur	396,800	377,900	18,900
Total Credits	513,268	488,849	24,419
Net Annualized Cost	679,870	688,810	(8,940)

Table 6-9

ANNUALIZED COST SUMMARY
 AMINE TREATMENT WITH CLAUS SULFUR RECOVERY
EXPANSION OF 10.16 Mg/D (10 LT/D) PLANT TO 10.668 Mg/D (10.5 LT/D)
 (January 1985 Dollars)

10% H₂S

	10.66 Mg/D	10.16 Mg/D	Incremental
Direct Operating Cost			
Operating labor	210,000	210,000	0
Supervision	60,900	60,900	0
Maintenance, repairs	83,000	82,070	930
Supplies, laboratory	16,800	16,800	0
Utilities	172,300	164,100	8,200
Chemicals, catalyst	7,268	6,922	346
Indirect Operating Cost			
Overhead (50 percent of all labor costs)	155,300	155,300	0
Capital Recovery (10 percent rate of return, 15 years equipment life)	372,580	368,070	4,510
Taxes (one percent of capital cost)	28,330	27,990	340
Insurance (one percent of capital cost)	28,330	27,990	340
Administration (two percent of capital cost)	56,660	55,980	680
Total Annualized Cost Without Credit	1,191,470	1,176,120	15,350
Credits:			
4300 Kp steam	0	0	0
1760 Kp steam	100,100	95,360	4,740
352 Kp steam	0	0	0
106 Kp steam	6,601	6,287	314
Steam Condensate	9,767	9,302	465
Sulfur	396,800	377,900	18,900
Total Credits	513,268	488,849	24,419
Net Annualized Cost	678,200	687,270	(9,070)

Direct and indirect installation factors from Reference 7 were then applied to each model plant purchased equipment cost to obtain total installed cost.

The capital costs for the four model plants are presented in Table 6-10.

6.3.2 Annualized Costs

The annualized costs for the LO-CAT® model plants have also been estimated. Most of the direct operating costs were taken directly from Reference 3, interpolating values from tabular data therein. Operating labor was calculated based on one-third of a man per shift for all model plants (Reference 5), 350 days per year, and a labor rate of \$15 per hour (see Table 6-2). Supervision was figured at 15 percent of labor and maintenance labor was assumed to equal operating labor (Reference 6). Catalyst costs, obtained from Reference 7, were found to vary according to the method used to separate the recovered sulfur from the catalyst solution.

Indirect costs were factored from capital costs or direct operating costs using factors from References 6 and 8. As with the amine-Claus control system, capital recovery is based on a 10 percent rate of return and a 15-year equipment life. There are no waste disposal costs for spent catalyst, in accordance with Reference 3. Product recovery credits for sulfur have also been calculated. Note that because the LO-CAT® process directly produces the desulfurized refinery fuel gas, these model plants are designed to meet (or better) the limit of 162 parts per million (ppm) H₂S in the gas. The design sulfur removal efficiencies vary with H₂S inlet concentration, with a consequent variation in sulfur recovery. Note also that there is some question about the quality of sulfur recovered from the LO-CAT® process. To be conservative, therefore, each annualized cost table includes a cost for sulfur disposal, calculated at a unit cost of \$24.60 per long ton (\$24.22 per Mg). The total annualized

Table 6-10

CAPITAL COST SUMMARY
LO-CAT® PROCESS
 (January 1985 Dollars)

	Cost in Thousands of Dollars ^a Plant Capacity, Mg/D (LT/D) Sulfur			
	<u>1.016(1)</u>	<u>2.032(2)</u>	<u>5.08(5)</u>	<u>10.16(10)</u>
Purchased equipment ^b (A)	520	810	1460	2380
Taxes and freight(0.08 ^c XA)	<u>42</u>	<u>65</u>	<u>117</u>	<u>190</u>
Base Cost (B)	562	875	1577	2570
Direct/indirect installation	281	438	789	1285
(0.50 ^d XB)	—	—	—	—
Total Installed Cost	843	1313	2366	3855

a Cost for each capacity applies for inlet hydrogen sulfide concentrations of 1, 5, or 10 percent.

b Costs were developed from information in Reference 3 and include instrumentation.

c Reference 6 was the source of this factor.

d Overall installation factor suggested by Reference 7 for refinery installations.

cost for the process is therefore bracketed between a low figure that includes full credit and a high figure that includes the full cost of disposal. In reality, the annualized cost would probably fall in between, depending on the quality of the sulfur produced and the available markets for sulfur of that quality.

The annualized costs for the model plants are presented in Tables 6-11 through 6-14.

6.4 COST-EFFECTIVENESS

The cost-effectiveness values are determined from the costs developed in Sections 6.2 and 6.3 and the emission reductions presented in Chapter 5. The cost-effectiveness values represent the estimated cost (in January 1985 dollars) that would be incurred by a refinery for each ton of SO₂ emissions that is controlled.

Cost-effectiveness values for both control systems and all model plants are shown in Tables 6-15 and 6-16, for the amine-Claus and LO-CAT® processes, respectively. All cost-effectiveness values are based on a baseline of zero control. For both processes, the C/E ratio increases as the model plant size decreases.

The amine-Claus system's cost-effectiveness improves to a credit of \$10 to \$12/Mg at the 101.6 Mg/D model plant size. The model plant size at where the transition from cost to credit occurs is approximately 30 Mg/D. (Note that because costs for the 203.2 and 508 Mg/D plants are multiples of the cost for the 101.6 Mg/D plant, the cost-effectiveness for all plants 101.6 Mg/D and larger is exactly equal in this analysis.) Cost-effectiveness is not a strong function of inlet H₂S concentration in the range of concentrations examined in this analysis.

Table 6-11

ANNUALIZED COST SUMMARY^{a,b}
 LO-CAT® PROCESS
 1.016 Mg/D (1 LT/D) PLANT
 (January 1985 Dollars)

	Cost in Thousands of Dollars
Direct Operating Cost	
Operating labor	42
Supervision	6
Maintenance labor	42
Utilities (electricity)	2
Catalysts and chemicals	64
Indirect Operating Cost	
Overhead (50 percent of all labor costs)	45
Capital Recovery (10 percent rate of return, 15 years equipment life)	111
Taxes (one percent of capital cost)	8
Insurance (one percent of capital cost)	8
Administration (two percent of capital cost)	17
Total Annualized Cost Without Credit	346
Sulfur Credit ^c	38
Net Annualized Cost (With Credit)	<u>308</u>
Sulfur Disposal ^c	8
Total Annualized Cost with Disposal	<u>355</u>

^aWith the exception of the sulfur credit and sulfur disposal cost, the costs shown apply for inlet H₂S concentration of 1, 5, and 10 percent. See Note c.

^bTotals may not add due to rounding.

^cThe sulfur credit and disposal cost shown are for 1 percent H₂S. For other inlet H₂S concentrations, the sulfur credit would be as follows: 5 percent H₂S -- 39; 10 percent H₂S -- 39. The disposal cost would be: 5 percent H₂S -- 9; 10 percent H₂S -- 9.

Table 6-12

ANNUALIZED COST SUMMARY^{a,b}
 LO-CAT® /PROCESS
 2.032 Mg/D (2 LT/D) PLANT
 (January 1985 Dollars)

	Cost in Thousands of Dollars
Direct Operating Cost	
Operating labor	42
Supervision	6
Maintenance labor	42
Utilities (electricity)	3
Catalysts and chemicals	129
Indirect Operating Cost	
Overhead (50 percent of all labor costs)	45
Capital Recovery (10 percent rate of return, 15 years equipment life)	173
Taxes (one percent of capital cost)	13
Insurance (one percent of capital cost)	13
Administration (two percent of capital cost)	26
Total Annualized Cost Without Credit	493
Sulfur Credit ^c	77
Net Annualized Cost (With Credit)	<u>416</u>
Sulfur Disposal ^c	17
Total Annualized Cost with Disposal	<u>509</u>

^aWith the exception of the sulfur credit and sulfur disposal cost, the costs shown apply for inlet H₂S concentration of 1, 5, and 10 percent. See Note c.

^bTotals may not add due to rounding.

^cThe sulfur credit and disposal cost shown are for 1 percent H₂S. For other inlet H₂S concentrations, the sulfur credit would be as follows: 5 percent H₂S -- 78; 10 percent H₂S -- 78. The disposal cost would be: 5 percent H₂S -- 17; 10 percent H₂S -- 17.

Table 6-13

ANNUALIZED COST SUMMARY^{a,b}
 LO-CAT® PROCESS
 5.08 Mg/D (5 LT/D) PLANT
 (January 1985 Dollars)

	Cost in Thousands of Dollars
Direct Operating Cost	
Operating labor	42
Supervision	6
Maintenance labor	42
Utilities (electricity)	5
Catalysts and chemicals	142
Indirect Operating Cost	
Overhead (50 percent of all labor costs)	45
Capital Recovery (10 percent rate of return, 15 years equipment life)	311
Taxes (one percent of capital cost)	24
Insurance (one percent of capital cost)	24
Administration (two percent of capital cost)	47
Total Annualized Cost Without Credit	688
Sulfur Credit ^c	192
Net Annualized Cost (With Credit)	<u>496</u>
Sulfur Disposal ^c	42
Total Annualized Cost with Disposal	<u>731</u>

^aWith the exception of the sulfur credit and sulfur disposal cost, the costs shown apply for inlet H₂S concentration of 1, 5, and 10 percent. See Note c.

^bTotals may not add due to rounding.

^cThe sulfur credit and disposal cost shown are for 1 percent H₂S. For other inlet H₂S concentrations, the sulfur credit would be as follows: 5 percent H₂S -- 195; 10 percent H₂S -- 195. The disposal cost would be: 5 percent H₂S -- 43; 10 percent H₂S -- 43.

Table 6-14

ANNUALIZED COST SUMMARY^{a,b}
 LO-CAT[®] PROCESS
 10.16 Mg/D (10 LT/D) PLANT
 (January 1985 Dollars)

	Cost in Thousands of Dollars
Direct Operating Cost	
Operating labor	42
Supervision	6
Maintenance labor	42
Utilities (electricity)	7
Catalysts and chemicals	284
Indirect Operating Cost	
Overhead (50 percent of all labor costs)	45
Capital Recovery (10 percent rate of return, 15 years equipment life)	507
Taxes (one percent of capital cost)	39
Insurance (one percent of capital cost)	39
Administration (two percent of capital cost)	77
Total Annualized Cost Without Credit	1088
Sulfur Credit ^c	385
Net Annualized Cost (With Credit)	<u>703</u>
Sulfur Disposal ^c	85
Total Annualized Cost with Disposal	<u>1172</u>

^aWith the exception of the sulfur credit and sulfur disposal cost, the costs shown apply for inlet H₂S concentration of 1, 5, and 10 percent. See Note c.

^bTotals may not add due to rounding.

^cThe sulfur credit and disposal cost shown are for 1 percent H₂S. For other inlet H₂S concentrations, the sulfur credit would be as follows: 5 percent H₂S -- 390; 10 percent H₂S -- 391. The disposal cost would be: 5 percent H₂S -- 86; 10 percent H₂S -- 86.

Table 6-15

COST EFFECTIVENESS
SULFUR DIOXIDE CONTROL
AMINE TREATMENT WITH CLAUS SULFUR RECOVERY^a

(January 1985 Dollars)

Inlet Concentration	Plant Size Mg/day (LT/D)	Annualized Cost (Credit) \$10 ³ /yr	Mg/yr	SO ₂ Removed tons/yr	\$/Mg	C/E \$/ton
1% H ₂ S	5.08 (5)	762	3,435	3,787	222	201
	10.16 (10)	695	6,870	7,573	101	92
	50.8 (50)	(11)	34,350	37,867	(0.32)	(0.29)
	101.6 (100)	(702)	68,700	75,734	(10.2)	(9.3)
	0.51 (0.50) ^b	(8.7)	344	379	(25.3)	(23.0)
5% H ₂ S	5.08 (5)	759	3,435	3,787	221	200
	10.16 (10)	688	6,870	7,573	100	91
	50.8 (50)	(47)	34,350	37,867	(1.37)	(1.24)
	101.6 (100)	(775)	68,700	75,734	(11.3)	(10.2)
	0.51 (0.50) ^b	(8.9)	344	379	(26.0)	(23.6)
10% H ₂ S	5.08 (5)	758	3,435	3,787	221	200
	10.16 (10)	686	6,870	7,573	100	91
	50.8 (50)	(55)	34,350	37,867	(1.60)	(1.45)
	101.6 (100)	(789)	68,700	75,734	(11.5)	(10.4)
	0.51 (0.50) ^b	(9.1)	344	379	(26.4)	(23.9)

^a The cost-effectiveness ratios for 200 LT/D and 500 LT/D plants would be the same as shown for 100 LT/D, in that the annualized costs for these plants are multiples of the 100 LT/D costs in this analysis. However, if large plants were to be custom-designed, their C/E ratios would be expected to be lower, reflecting economies of scale in the capital costs.

^b Expansion case (10.16 to 10.668 Mg/D). All values shown on these lines are incremental between the base and expansion capacities.

Table 6-16

COST EFFECTIVENESS
SULFUR DIOXIDE CONTROL
LO-CAT® PROCESS

(January 1985 Dollars)

Inlet Concentration	Plant Size Mg/D (LT/D)	Sulfur Status	Annualized Cost (Credit) \$10 ³ /yr*	SO ₂ Removed Mg/yr	tons/yr	C/E	
						\$/Mg	\$/ton
1% H ₂ S	1.016 (1)	recovered	308	700	771	440	399
		disposed	355	700	771	507	460
	2.032 (2)	recovered	416	1,399	1,543	297	270
		disposed	509	1,399	1,543	364	330
	5.08 (5)	recovered	496	3,499	3,856	142	129
		disposed	731	3,499	3,856	209	189
	10.16 (10)	recovered	703	6,997	7,713	100	91
		disposed	1,172	6,997	7,713	168	152
5% H ₂ S	1.016 (1)	recovered	308	709	781	434	394
		disposed	355	709	781	501	455
	2.032 (2)	recovered	415	1,418	1,563	293	266
		disposed	509	1,418	1,563	359	326
	5.08 (5)	recovered	493	3,545	3,907	139	126
		disposed	732	3,545	3,907	206	187
	10.16 (10)	recovered	698	7,090	7,815	98	89
		disposed	1,173	7,090	7,815	165	150
10% H ₂ S	1.016 (1)	recovered	308	710	783	434	393
		disposed	355	710	783	500	453
	2.032 (2)	recovered	415	1,420	1,565	292	265
		disposed	509	1,420	1,565	358	325
	5.08 (5)	recovered	493	3,551	3,914	139	126
		disposed	732	3,551	3,914	206	187
	10.16 (10)	recovered	697	7,101	7,827	98	89
		disposed	1,173	7,101	7,827	165	150

*For the special case where neither a sulfur recovery credit nor a disposal cost were taken, the annualized cost would be found in Table 6-11, -12, -13, or -14. The cost-effectiveness would then be the quotient of this cost (the "Total Annualized Cost Without Credit") and the "SO₂ Removed" value in column 5.

The most interesting aspect of Tables 6-15 and 6-16 is the comparison of cost-effectiveness between the two control systems at the common model plant sizes--5.08 and 10.16 Mg/D. At the 10.16 Mg/D size, the cost-effectiveness ratios of the amine-Claus and LO-CAT® systems (sulfur recovery case) are essentially equal. But as the model plant size decreases to 5.08 Mg/D, the amine-Claus system becomes less cost-effective than the LO-CAT® system (sulfur recovery case). This indicates that for these lower sulfur loadings, the LO-CAT® would be the preferred control system to use.

6.5 COST COMPARISON

To determine their representativeness, the capital costs plus the operating and the maintenance (O&M) costs in Sections 6.2 and 6.3 were compared to cost data obtained from refineries that responded to Section 114 letter requests. Of the 13 plants covered in these 114 responses, seven contained enough cost/process data to allow meaningful comparisons. Because some of the respondents requested their cost data to be held confidential, no specific results are shown herein. However, some general information about the cost comparisons can be provided:

- o After escalating all costs to January 1985 dollars, costs were compared for seven amine treating units (ATU's), one LO-CAT® system, and one combination ATU-CLAUS system. (Note: some refineries provided data on more than one system.)
- o All but one of the capital cost sets differed by less than $\pm 50\%$, the nominal accuracy of the total installed costs in the chapter.
- o Larger discrepancies were seen in the O&M costs, due to differences in unit prices, cost allocations, accounting methods, and other factors. In all but one case, the O&M costs in the chapter were higher than the respondents'.

- o Details of the cost comparison have been placed in the confidential portion of the project docket.

6.6 REFERENCES FOR CHAPTER 6

1. Shumaker, Jeffrey L., Memorandum to Docket A-80-20A, dated September 11, 1984, concerning sweetening plant cost factors.
2. Review of New Source Performance Standards for Petroleum Refinery Claus Sulfur Recovery Plants, U.S. Environmental Protection Agency, EPA-450/3-83-014, August 1983.
3. Hardison, L.C., "Minimizing Gas Treating Costs with LO-CAT® for H₂S Removal", for presentation at the Panhandle Plains Regional Meeting of the Gas Processors Association, Amarillo, TX, October 11, 1984.
4. Hardison, L.C., "Go from H₂S to S in One Unit", Hydrocarbon Processing, April 1985, pp. 70-71.
5. Telecons--Thomas Beggs (JACA Corporation, Fort Washington, PA) with Mr. L.C. Hardison (President, ARI Technologies, Inc., Palatine, IL): April 1, 1985; May 21, 1985; and June 10, 1985.
6. Vatavuk, William M. and Neveril, Robert B., "Estimating Air-Pollution Control Costs--Part II: Estimating Capital and Operating Costs," Chemical Engineering, November 3, 1980, pp. 157-162.
7. Letter from L. C. Hardison (ARI Technologies, Inc., Palatine, IL) to W. L. Elmore (U.S. Environmental Protection Agency, Research Triangle Park, NC), February 26, 1986.
8. Peters, M.S. and Timmerhaus, K.D., Plant Design and Economics for Chemical Engineers, Third Edition. McGraw-Hill, New York, NY, 1980, pp. 203-209.

APPENDIX A

Listed below are the cost factors that were obtained or derived from Reference 1 and used in estimating capital and annualized costs for the amine treating process.

<u>Component</u>	<u>Cost Factor</u>
Capital cost (10% H ₂ S)	\$0.51 per standard cubic feet per day (SCFD) of acid gas*
Capital cost (1% H ₂ S or 5% H ₂ S)	0.9 (cap. cost at 10%) +0.1 (cap. cost of 10%) $\left[\frac{10}{\text{H}_2\text{S conc.}} \right]^{0.6}$
Operating labor (including some maintenance)	\$126,000 per year
Maintenance	\$0.01 per year per SCFD acid gas
Utilities (heat and electricity)	\$0.274 per year per SCFD acid gas
Chemicals	\$0.0187 per year per SCFD acid gas

* The acid gas flowrate is related to the sulfur loading, as follows:

$$\text{Flowrate (SCFD)} = 33,160 \times S$$

where: S = sulfur fed to amine treater (LT/D)

Via this equation, the above factors may be rewritten in terms of the sulfur loading.

APPENDIX B

This section presents three examples of line item annualized costs for the control systems analyzed. The model plant used for illustration is the 5.08 Mg/D (5 LT/D) plant with an inlet concentration of one percent H_2S . The examples are shown in Tables B-1, B-2, and B-3.

Table B-1

LINE ITEM ANNUALIZED COST EXAMPLE^a
AMINE TREATMENT, 5 LT/D, 1% H₂S
 (January 1985 Dollars)

Direct			\$126,000
Operating labor ^b	(direct from Reference 1)		18,900
Supervision	15% of above: $\$126,000 \times 0.15$		1,658
Maintenance	\$0.01 per SCFD: $165,800 \times 0.01$		45,429
Utilities	\$0.274 per SCFD: $165,800 \times 0.274$		3,100
Chemicals	\$0.0187 per SCFD: $165,800 \times 0.0187$		
Indirect			
Overhead	50% of labor: $(126,000 + 18,900) \times 0.50$	72,450	
Capital recovery	CRF (10 percent, 15 years) = 0.1315: $\$109,925 \times 0.1315$	14,455	
Taxes	1 percent of capital cost: $\$109,925 \times 0.01$	1,099	
Insurance	1 percent of capital cost: $\$109,925 \times 0.01$	1,099	
Administration	2 percent of capital cost: $\$109,925 \times 0.02$	<u>2,199</u>	
			<u>\$286,389</u>

^a Direct factors from Reference 1. (See Appendix A).

^b Includes an unspecified amount of maintenance labor.

Table B-2

LINE ITEM ANNUALIZED COST EXAMPLE^a
CLAUS SULFUR RECOVERY PLANT 5 LT/D, 1% H₂S
 (January 1985 Dollars)

Direct

Operating Labor	$\frac{5600 \text{ hours}}{\text{year}} \times \frac{\$15}{\text{hour}}$	\$84,000
Supervision	$\frac{2100 \text{ hours}}{\text{year}} \times \frac{\$20}{\text{hour}}$	42,000
Maintenance, repair ^b	3 percent of capital cost: $\$2,285,000 \times 0.03$	68,550
Supplies, lab	20 percent of operating labor: $\$84,000 \times 0.20$	16,800
4300 Kp steam	$\frac{200 \text{ Mg}}{\text{Year}} \times \frac{\$17.50}{\text{Mg}}$	3,500
Boiler feedwater	$\frac{3267 \text{ Mg}}{\text{Year}} \times \frac{\$3.60}{\text{Mg}}$	11,760
Electricity	$\frac{764 \text{ Gj}}{\text{Year}} \times \frac{\$15.28}{\text{Gj}}$	11,670
Fuel gas	$\frac{2423 \text{ Gj}}{\text{Year}} \times \frac{\$4}{\text{Gj}}$	9,690
Catalyst	$\frac{\$655}{10 \text{ LT/D}} \times 5 \text{ LT/D} \times 1.1$	361

Indirect

Overhead	50 percent of labor: $0.50 \times (84,000 + 42,000 + 0.5 \times 68,550)$	80,138
Capital recovery	CRF (10 percent, 15 years) = $0.1315: \$2,285,000 \times 0.1315$	300,478
Taxes	1 percent of capital cost: $\$2,285,000 \times 0.01$	22,850
Insurance	1 percent of capital cost: $\$2,285,000 \times 0.01$	22,850

Table B-2 (cont'd)

Indirect (cont'd)

Administration	2 percent of capital cost: \$2,285,000 X 0.02	45,700
Total Annualized Cost w/o Credit		720,347

Credits

1760 Kp steam	$\frac{2934 \text{ Mg}}{\text{year}} \times \frac{\$16.25}{\text{Mg}}$	47,680
106 Kp steam	$\frac{286 \text{ Mg}}{\text{year}} \times \frac{\$11.00}{\text{Mg}}$	3,144
Steam condensate	$\frac{1550 \text{ Mg}}{\text{year}} \times \frac{\$3.00}{\text{Mg}}$	4,651
Sulfur	$\frac{5.08 \text{ Mg}}{\text{day}} \times 0.966 \text{ recovery} \times \frac{350 \text{ days}}{\text{year}} \times \frac{\$110}{\text{Mg}}$	188,930
Total Credits		244,405

NET ANNUALIZED COST

\$475,942

a Consumption figures derived from cost data in Reference 2.

b Assumed to be equal parts labor and materials.

Table B-3

LINE ITEM ANNUALIZED COST EXAMPLE^a
LO-CAT® PROCESS, 5 LT/D, 1% H₂S

Direct

Operating Labor	$\frac{8 \text{ hours}}{\text{day}} \times \frac{350 \text{ days}}{\text{year}} \times \frac{\$15}{\text{hour}}$	\$42,000
Supervision	15% of above: $\$42,000 \times 0.15$	6,300
Maintenance labor	Same as operating	42,000
Utilities ^b	$\frac{\$2.70}{\text{LT}} \times \frac{5 \text{ LT}}{\text{day}} \times \frac{350 \text{ days}}{\text{year}}$	4,725
Catalysts and Chemicals ^b	$\frac{\$81.20}{\text{LT}} \times \frac{5 \text{ LT}}{\text{day}} \times \frac{350 \text{ days}}{\text{year}}$	142,100

Indirect

Overhead	50 percent of labor: $(42,000 + 6,300 + 42,000) \times 0.50$	45,200
Capital recovery	CRF (10 percent, 15 years) = 0.1315: $\$2,366,000 \times 0.1315$	311,100
Taxes	1 percent of capital cost: $\$2,366,000 \times 0.01$	23,700
Insurance	1 percent of capital cost: $\$2,366,000 \times 0.01$	23,700
Administration	2 percent of capital cost: $\$2,366,000 \times 0.02$	47,300
Total Annualized Cost w/o Credit		688,100
Sulfur Credit	$\frac{5.08 \text{ Mg}}{\text{day}} \times 0.9838 \text{ recovery} \times \frac{350 \text{ days}}{\text{year}} \times \frac{\$110}{\text{Mg}}$	192,400
NET ANNUALIZED COST (WITH CREDIT)		\$495,700
Sulfur Disposal	$\frac{5.08 \text{ Mg}}{\text{day}} \times 0.9838 \text{ recovery} \times \frac{350 \text{ days}}{\text{year}} \times \frac{\$24.22}{\text{Mg}}$	42,400
TOTAL ANNUALIZED COST WITH DISPOSAL		\$730,500

^a In most cases, unit consumption data were taken from References 3, 4, and 5.

^b Unit costs for utilities and catalysts are not the same for all plant sizes, as the quantities required of steam, electricity, catalyst, etc., vary nonlinearly with capacity.